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PROCEEDINGS: EPA'S INDUSTRY BRIEFING ON THE ORGANIC-ACID-ENHANCED LIMESTONE FGD PROCESS

(JULY 1984)

J. David Mobley, Briefing Chairman

Air and Energy Engineering Research Laboratory
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Research Triangle Park, North Carolina 27711

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16. ABSTRACT The proceedings document presentations made during an EPA-sponsored industry briefing, July 19, 1984, in San Antonio, TX. The briefing dealt with the status of EPA's research activities on the organic-acid-enhanced limestone flue gas desulfurization (FGD) process. Subjects covered included: (1) a technical and economic evaluation of organic acid addition to the San Miguel FGD system--results of parametric tests and the utility's perspective, (2) results of the first 2 years of commercial operation of an organic-acid-enhanced FGD system, (3) economics of retrofitting Big Rivers Electric Corporation's lime-based FGD system to organic-acid-enhanced limestone operations, and (4) an economic evaluation of limestone and lime FGD processes for new systems. The briefing provided users, architects, engineers, vendors, consultants, and government personnel with a comprehensive assessment of this innovative technology for controlling SO2 emissions.

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This document has been reviewed in accordance with U.S. Environmental Protection Agency policy and approved for publication. Mention of trade names or commercial products does not constitute endorsement or recommendation for use.

PREFACE

EPA's Industrial Environmental Research Laboratory* in Research Triangle Park, NC (IERL-RTP) periodically sponsors symposia and industry briefings for the transfer of information regarding research, development, and application activities with the objective of further accelerating the development and commercialization of technology. One of the major IERL-RTP efforts for the past several years has been advancement of the technology for flue gas desulfurization (FGD). A key element of the FGD program has been the advancement of lime/limestone wet scrubbing technology, which has led to the development of the organic-acid-enhanced limestone FGD process.

The July 19, 1984 Industry Briefing was designed to provide interested people with EPA's latest findings on the use of organic acid in lime/limestone scrubbers. The presentations covered testing work performed at the laboratory, bench, pilot, prototype, industrial, and utility scales. In addition, an economic assessment of the use of the technology on new and retrofit systems was included. The presentations were complemented by a tour of the FGD system at San Miguel Electric Cooperative's Jourdanton Station in Jourdanton, Texas; a summary of the tour is provided as an appendix to these proceedings.

More than 220 people representing electric utilities, industrial sources, architect and engineering firms, vendors, consultants, and governmental personnel attended the Conference. The Chairman of the Briefing was J. David Mobley, an Environmental Engineer with the Industrial Environmental Research Laboratory.

(*) Recently redesignated the Air and Energy Engineering Research Laboratory.

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ABSTRACT

The proceedings document presentations made during an EPA-sponsored industry briefing which was held on July 19, 1984 in San Antonio, Texas. The briefing dealt with the status of EPA's research activities on the organic-acid-enhanced limestone flue gas desulfurization process. Subjects covered included: (1) technical and economic evaluation of organic acid addition to the San Miguel FGD system - results of parametric tests, (2) technical and economic evaluation of organic acid addition to the San Miguel FGD system - the utility's perspective, (3) results of the first two years of commercial operation of an organic-acid-enhanced FGD system, (4) economics of retrofitting Big Rivers Electric Corporation's lime based FGD system to organic-acid-enhanced limestone operations, and (5) an economic evaluation of limestone and lime flue gas desulfurization processes for new systems. The briefing provided users, architects and engineers, vendors, consultants, and government personnel with a comprehensive assessment of this innovative technology for controlling sulfur dioxide emissions.

ACKNOWLEDGMENTS

The Chairman of the Briefing would like to acknowledge the support of the authors and speakers at the Briefing. In addition, the support of San Miguel Electric Cooperative is gratefully acknowledged for hosting the tour of the flue gas desulfurization system at their Jourdanton Station. Robert P. Metcalfe and Robert Cmiel contributed significantly to the success of this endeavor. Personnel of Radian Corporation were also very instrumental in the organization and conduct of the Briefing. In particular, the contributions of James C. Dickerman, Jack M. Burke, Holly G. Rossi, and Ann E. Behl were especially significant.

AGENDA

EPA's Industry Briefing on the
Organic-Acid-Enhanced Limestone FGD Process

San Antonio, Texas
July 19, 1984

6:30 - 8:00	Registration
8:00 - 12:00	Tour of the FGD System at San Miguel Electric Cooperative's Jourdanton Station
12:00 - 1:00	Luncheon
1:00 - 1:15	Introductory Remarks J. David Mobley (EPA)
1:15 - 2:15	Technical and Economic Evaluation of Organic Acid Addition to the San Miguel FGD System Part 1 - Results of Parametric Tests Jack M. Burke (Radian) Part 2 - The Utility's Perspective Robert P. Metcalfe and Robert Cmiel (San Miguel Electric Cooperative)
2:15 - 2:35	Results of the First Two Years of Commercial Operation of an Organic-Acid-Enhanced FGD System N. Dale Hicks (Springfield City Utilities)
2:35 - 2:55	Economics of Retrofitting Big Rivers Electric Corporation's Lime Based FGD System to Organic-Acid-Enhanced Limestone Operations Dennis Laslo (Peabody Process Systems)
2:55 - 3:15	Economic Evaluation of Limestone and Lime Flue Gas Desulfurization Processes for New Systems Robert L. Torstrick (Tennessee Valley Authority)
3:15 - 3:30	Briefing Summary J. David Mobley (EPA)
3:30	Adjournment

TECHNICAL AND ECONOMIC EVALUATION OF ORGANIC ACID ADDITION TO THE SAN MIGUEL FGD SYSTEM

PART 1 - RESULTS OF PARAMETRIC TESTS

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Presented at:

EPA's Industry Briefing on the
Organic-Acid-Enhanced Limestone FGD Process
San Antonio, TX

July 19, 1984

ABSTRACT

This paper summarizes the results of organic acid addition tests at a commercial FGD system. The tests were conducted at San Miguel Electric Cooperative's 410 MW lignite-fired Unit 1 located outside Jourdanton, Texas. During the program, several organic acid mixtures were tested over a range of operating conditions. The objective of these tests was to determine if the use of organic acids would allow San Miguel to reduce FGD system operating costs. Based on the test results, a cost analysis indicated that the use of organic acid addition at San Miguel will result in a first-year cost savings of over $600,000. In terms of cumulative net present worth, the estimated savings over a fifteen-year period will be $7.2 million in 1984 dollars.

TECHNICAL AND ECONOMIC EVALUATION OF ORGANIC ACID ADDITION TO THE SAN MIGUEL FGD SYSTEM

Part 1 - Results of Parametric Tests

This paper presents the results of a test program which characterized the technical and economic aspects of organic acid addition to a commercial FGD system. The program was sponsored by EPA's Industrial Environmental Research Laboratory at Research Triangle Park (IERL-RTP) in cooperation with San Miguel Electric Cooperative, Inc. Radian Corporation was responsible for providing San Miguel with engineering and analytical support during the test program. This paper presents a summary of the results of an engineering analysis of the organic acid addition tests conducted at the San Miguel plant and a site-specific economic analysis which evaluates the cumulative net present worth of the organic acid addition options tested during the program.

BACKGROUND

Many of the existing and planned FGD systems in the United States use limestone as an alkaline reagent to react with SO_2 and produce a calcium sulfite and/or sulfate waste sludge. Although this technology has been under development for nearly 15 years, many of the existing limestone systems have problems in meeting design levels of reagent utilization* and SO_2 removal. One common problem is an inability to meet SO_2 emission regulations at low stoichiometric ratios* (i.e., 1.1 or less). Often limestone FGD systems are forced to operate at stoichiometric ratios of 1.3 or more.

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*Reagent utilization and stoichiometric ratio (S.R.) are both measures of the quantity of limestone used relative to the SO_2 removed in an FGD system. In this paper, S.R. = (moles $CaCO_3$ added) ÷ (moles SO_2 removed), and reagent utilization = 100 percent ÷ (S.R.).

Not only does this result in increased operating costs, but it can lead to other problems (e.g., mist eliminator pluggage) which may make it difficult to operate the FGD system. In some cases even extremely high limestone stoichiometries (1.5 or greater) will not allow a utility to meet emission limits and, depending on specific SO_2 emission regulations, the boiler may have to be derated because of the inability of the FGD system to perform at design levels.

Most problems with poor SO_2 removal or poor limestone utilization are symptoms of poor mass transfer (both gas/liquid and solid/liquid) in the FGD system. Many of these problems can be corrected with equipment modifications but, for an operating FGD system, such modifications can prove quite costly. Another alternative is the use of an organic acid buffer to improve the mass transfer rate of SO_2 from the flue gas into the scrubber slurry.

Organic acids were identified as potentially attractive buffering agents for FGD systems more than 10 years ago [1]. With a constant slurry feed pH, use of an organic acid which buffers in the correct range can help maintain a higher average slurry pH in the absorber of an FGD system and therefore increase SO_2 removal. Organic acid buffers can also be used to maintain constant SO_2 removal efficiency at a lower pH. For FGD systems which can meet SO_2 emission regulations only at very high absorber feed pHs (i.e., high limestone stoichiometries), an organic acid buffer can be used to meet SO_2 emission limits at a lower pH and a consequently lower stoichiometry. Depending on the specific situation, the costs of organic acid addition may be more than offset by reduced limestone consumption, and a net reduction in costs can result.

EPA's Industrial Environmental Research Laboratory is responsible for conducting research to improve the performance and reliability of lime and limestone FGD systems. Since 1977, EPA has studied the use of adipic acid* as a buffering agent for use in FGD systems. Initial tests with

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*Adipic acid is a dibasic organic acid which buffers in the 3 to 6 pH range.

adipic acid at EPA's laboratories and in a pilot unit at Tennessee Valley Authority's Shawnee Power Plant showed promising results. Based upon those initial tests, EPA sponsored a full-scale demonstration of adipic acid addition to a FGD system at Springfield City Utilities' Southwest Power Plant. As part of the Springfield test program, a mixture of dibasic acids* (consisting of adipic, glutaric, and succinic acids) was also tested as a buffering agent. The results of these tests led to the permanent installation of a dibasic acid (DBA) addition system at the Springfield Plant [2].

In another EPA sponsored study, the costs of DBA addition were estimated for several operating FGD systems including San Miguel Electric Cooperative's 410 MW lignite fired Unit 1 located outside Jourdanton, Texas. For the San Miguel plant, the results of the economic analysis indicated that over a 15-year period, the cost of using DBA could range from an increase of $657,000 to a savings of $10,778,000 in terms of net present worth. This range in costs was a result of uncertainties in the DBA consumption rate, the estimated improvement in limestone utilization, and the SO_2 removal which could be achieved with DBA addition [3]. Because of the potential savings associated with DBA addition, San Miguel agreed to participate with EPA in a program to gather data on the technical and economic aspects of DBA use at the San Miguel plant.

OBJECTIVES AND APPROACH

The overall goal of the San Miguel test program was to expand the technical and economic data base on the use of organic acid in full-scale FGD systems. The approach used to meet this objective involved testing of several organic acid additivies at San Miguel Electrical Cooperative's Unit 1, a 410 MW lignite fired boiler. The Unit 1 FGD system consists of four 33 percent capacity modules with one typically used as a spare. Each module includes a venturi-type quench tower for saturation of the flue gas and a tray-type absorber where most of the SO_2 removal occurs. During the test

*This dibasic acid mixture is a by-product of adipic acid manufacture.

program, the three on-line modules were all operated with organic acid addition. Test durations ranged from 2 days to 1 week, and for each test samples were collected to permit completion of DBA material balances around the FGD system and measurement of representative limestone ($CaCO_3$) utilizations in the system. Process data were collected to quantify FGD operation including the overall SO_2 emission rate, the flue gas bypass rate, and typical operating pHs in the system.

For each test, the general operating goals were to maintain the plant SO_2 emissions at or below 1.2 $lb/10^6$ Btu*, the absorber pHs at 5.8, the quencher pHs at 5.4, the gas flow through the modules at or below the design level of 500,000 acfm/module, and the organic acid buffer capacity at a relatively constant level. Of these goals, maintaining SO_2 emissions below 1.2 $lb/10^6$ Btu took precedence. At times this resulted in deviation from one or more of the other test goals.

PLANT DESCRIPTION

San Miguel Cooperative's Unit 1 is a 410 MW lignite fired boiler located in Atascosca County outside of Jourdanton, Texas. The plant is a mine-mouth facility firing a high-sulfur (about 2 percent), high ash (about 28 percent), low heating value (about 5600 Btu/lb) Texas lignite. Because of the poor fuel quality, the plant requires operation of an air quality control system consisting of an electrostatic precipitator for particulate collection, followed by a Babcock & Wilcox designed FGD system for SO_2 removal. The FGD system was designed to meet a Federal and State SO_2 emission limit of 1.2 $lb/10^6$ Btu which corresponds to about 85 percent SO_2 removal at 2700 ppm SO_2 inlet concentration.

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*It is EPA policy to report measurements in SI units. However, because of convention in the electric utility industry, this paper uses British engineering units. A table of conversion factors is attached.

The San Miguel Unit 1 FGD system consists of four quench-absorber modules, two 115-ft diameter thickeners, thickener underflow and overflow storage tanks, a limestone feed preparation system, and a solids dewatering and disposal system.

Figure 1 is a schematic of the FGD system which shows one of four identical absorber modules. Two boiler ID fans exhaust flue gas to a common plenum which supplies the scrubber modules. Any of the modules can be isolated from the flue gas path by closing inlet and outlet dampers. In addition, there is a flue gas bypass which can be used if SO_2 emissions drop below the 1.2 $lb/10^6$ Btu limit. Since flue gas exiting the scrubber is reheated, hot flue gas bypass reduces the load on the reheat system and, therefore, increases overall plant efficiency.

Flue gas which enters the absorbers (about 1,500,000 acfm) first passes down through a venturi-type quench tower where the gas is saturated with water and some SO_2 removal occurs. The gas then turns 180° and passes up through a perforated tray type absorber where it is contacted with a recirculating slurry of calcium sulfite/sulfate and limestone. There, the majority of the SO_2 removal occurs. The flue gas then leaves the absorbers and passes through a 2-stage Chevron type mist eliminator before entering a reheat chamber common to all four modules. The scrubbed flue gas is mixed with bypassed flue gas and with ambient air which has been heated to 400°F by steam coils. After the flue gas is reheated, it enters the stack and is discharged to the atmosphere.

TEST PROGRAM

San Miguel became interested in organic acid addition to the Unit 1 FGD system when an economic analysis showed that use of organic acids could result in a net reduction in FGD system operating costs. Prior to the use of DBA at the San Miguel plant, the FGD system operated with very high limestone stoichiometries (i.e., greater than 1.4) which resulted in high limestone reagent costs. San Miguel was also interested in DBA addition

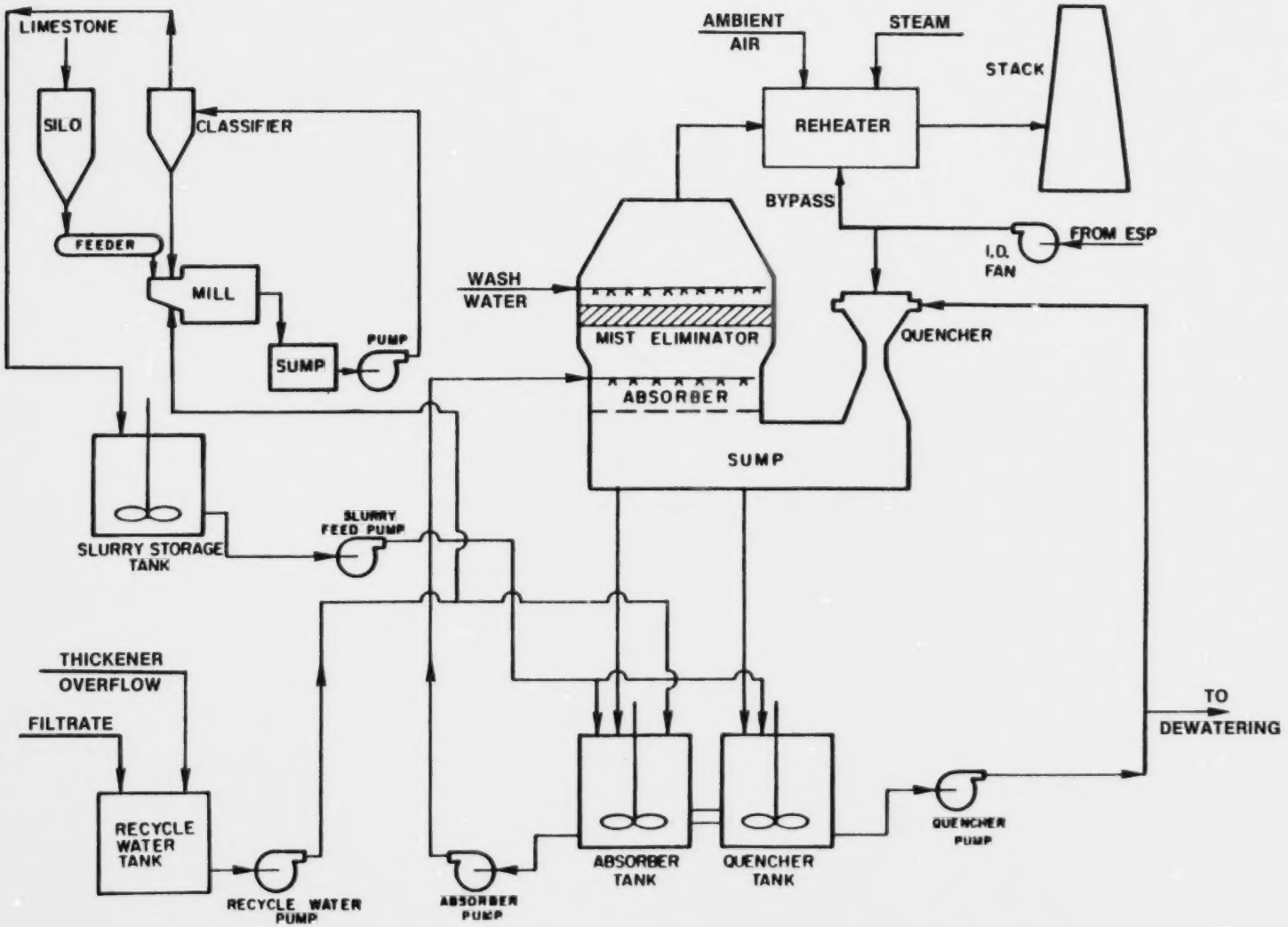

FIGURE 1 SAN MIGUEL FGD SYSTEM

because it offered the possibility of alleviating some of their operating problems, including mist eliminator pluggage.

The key factors affecting the economics of organic acid addition to the San Miguel FGD system were:

- The quantity of organic acid consumed in the FGD system at different organic acid concentrations in the circulating slurry,
- The limestone stoichiometry which could be achieved at different levels of organic acid addition to the system, and
- The quantity of flue gas which could be bypassed at different levels of organic acid addition to the system.

The initial test program for San Miguel was structured to address these key economic considerations.

Originally, the San Miguel test program was scheduled to evaluate two organic acid sources at three different concentrations in the recirculating slurry. The acid sources were:

- A DBA solution supplied by DuPont Petrochemicals of Victoria, Texas, and
- An organic acid solution supplied by Badische Corporation of Freeport, Texas.

The initial phase of the test program using the DuPont supplied DBA was carried out as scheduled. However, the tests using the Badische acid were terminated after the first test because of severe foaming in the absorber and quench recirculation tanks which made it very difficult to operate the FGD system. After this test, the program was modified to include tests not originally scheduled. The new tests included:

- Testing of an organic acid byproduct stream consisting primarily of maleic acid which was supplied by Badische Corporation, and

- Testing of manganese addition (as $MnSO_4$) to the FGD system to reduce organic acid consumption.

The San Miguel organic acid addition program consisted of nine tests including a baseline test prior to the start of DBA addition to the system. Table 1 presents the actual and planned operating conditions for each organic acid addition test. As shown, actual operating conditions (represented by test averages) differ slightly from the planned operating conditions.

The most significant difference between planned and actual test conditions was for the Badische EP-501 test where the buffer capacity was 25 percent of the planned level. This deviation is a result of several factors, including very high consumption rates measured for EP-501 and several late deliveries of the EP-501 material which caused the organic acid feed to be temporarily discontinued.

Table 2 presents typical compositions for the three organic acid mixtures tested at San Miguel. The compositions shown are from suppliers' analyses.

RESULTS

The discussion of the results of the San Miguel organic acid addition tests is organized into four areas:

- Performance results,
- Operation results,
- Additive effects, and
- Economic results.

TABLE 1. SAN MIGUEL ORGANIC ACID TEST CONDITIONS

Test ID	Organic Acid Tested	Buffer Capacity (ppm)		Absorber pH		Quench pH		SO_2 Emissions $lb/10^6$ Btu		Mn Concentration (ppm)	
		Planned	Actual	Planned	Actual	Planned	Actual	Planned	Actual	Planned	Actual
DuPont 1	DuPont: DBA	500	632	5.80	5.83	5.40	5.48	1.15	1.02	NA	1
DuPont 2	DuPont: DBA	1000	1415	5.80	5.77	5.40	5.47	1.15	0.98	NA	1
DuPont 3	DuPont: DBA	2000	2018	5.80	5.82	5.40	5.47	1.15	0.77	NA	1
Badische 1	Badische: EP-306	500	461	5.80	5.79	5.40	5.69	1.15	1.00	NA	1
Badische 2	Badische: EP-501	500	125	5.80	6.05	5.40	5.86	1.15	1.07	NA	1
Mn 1	DuPont: DBA	500	643	5.80	5.9	5.40	5.56	1.15	1.07	15	6
Mn 2	DuPont: DBA	500	574	5.80	5.9	5.40	5.57	1.15	1.02	30	10
Mn 3	DuPont: DBA	500	629	5.80	5.86	5.40	5.64	1.15	1.10	15	10

TABLE 2. TYPICAL COMPOSITIONS FOR THE ORGANIC ACID MIXTURES TESTED AT SAN MIGUEL

	Mixture		
	DuPont DBA	Badische EP-306	Badische EP-501
Wt. % acids as delivered	50	25	20
Component (% of total acids)			
Adipic	15	45	--
Succinic	25	3	--
Glutaric	60	5	--
Hydroxy Caproic	--	35	--
Maleic	--	--	80
Others	--	2	20

Each of these areas is summarized below.

Performance Results

The San Miguel test program examined different organic acids over a range of operation conditions. Table 3 summarizes many of the key measures of process performance for these tests. The data in Table 3 are useful in estimating the effects of various additives, and they can also be used to compare the San Miguel results with those of other programs. In terms of process costs, two of the most significant items included in Table 3 are organic acid consumption and limestone utilization. These are discussed in detail below.

Organic Acid Consumption--

Organic acid consumption is one of the most significant factors which can affect the costs of DBA addition to an FGD system. Previous tests have shown that organic acid consumption in an FGD system can be a function of several factors including buffer capacity* (which for a given acid or acid mixture is proportional to concentration), sulfite oxidation, acid composition, manganese concentration, pH, etc. For any of these factors that can be controlled, it is important to quantify how they affect organic acid consumption and also how they affect limestone utilization and bypass rate.

Figure 2 presents the organic acid consumption rates for the San Miguel tests as a function of buffer capacity. The figure also includes a curve which is a regression fit of the DuPont DBA test results. It should be noted that the consumption rates in Figure 2 represent the total organic acid loss rate in the San Miguel system including:

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*Buffer capacity is a relative measure of the ability of an organic acid or acid mixture to buffer over the pH range 6 to 3. For this study, buffer capacities are reported relative to the buffer capacity of adipic acid.

TABLE 3 SAN MIGUEL TEST SUMMARY

ITEM	TEST ID							
	DuPont 1	DuPont 2	DuPont 3	Badische 1	Badische 2	Mn 1	Mn 2	Mn 3
BUFFER CAPACITY (ppm)	632	1415	2018	461	125	643	574	629
BUFFER TYPE	DuPont DBA	DuPont DBA	DuPont DBA	EP-306	EP-501	DuPont DBA	DuPont DBA	DuPont DBA
MANGANESE CONC. (ppm)	1	1	1	1	1	6.9	10.8	10.8
DBA CONSUMPTION								
a. lb/hr	293	535	582	196	241	326	271	321
b. lb/ton SO2	18.0	34.4	35.2	12.9	17.1	28.7	24.4	29.7
INLET SO2 ($lb/10^6 Btu$)								
a. Coal Analysis	7.41	7.41	7.41	7.07	7.07	5.19	5.19	5.19
b. SO2 Monitors	7.68	6.45	7.55	7.20	6.48	6.08	5.70	5.59
SO2 EMISSIONS ($lb/10^6 Btu$)	1.02	0.98	0.77	1.00	1.07	1.07	1.02	1.10
SO2 REMOVAL - MODULES (%)								
a. Coal Analysis	91.1	94.1	95.8	87.5	87.8	84.0	86.0	83.6
b. SO2 Monitors	90.8	94.1	96.1	82.4	86.5	90.3	91.4	94.1
SO2 REMOVAL - SYSTEM (%)								
a. Coal Analysis	86.2	86.8	89.7	85.9	84.9	79.4	80.4	78.8
b. SO2 Monitors	86.7	84.8	89.9	86.1	83.5	82.4	82.1	80.3
SO2 REMOVED (tons/hr)	16.3	15.6	16.6	15.2	14.1	11.5	11.2	11.0
LIMESTONE UTILIZATION (%)								
a. Absorbers	85.9	86.3	74.0	65.7	79.1	79.3	82.1	82.3
b. Quenchers	85.0	89.4	91.3	82.7	84.8	87.8	92.6	88.4
OPERATING pHs								
a. Absorbers	5.83	5.77	5.82	5.79	6.05	5.90	5.90	5.86
b. Quenchers	5.48	5.47	5.47	5.69	5.86	5.56	5.57	5.64
GAS FLOWS (wet mscfm)								
a. Modules	1296	1266	1268	1358	1301	1295	1295	1300
b. Bypass	73.5	106.9	86.7	26.5	29.5	75.3	91.4	78.4
c. Reheat	109.9	57.1	192.0	182.7	192.5	143.0	131.7	167.1
MAXIMUM BYPASS (wet mscfm)								
a. Coal Analysis	89.4	130.5	150.5	48.5	50.9	84.6	115.8	78.9
b. SO2 Monitors	77.3	162.7	150.4	-39.0	53.5	127.4	162.2	201.9
SULFITE OXIDATION (%)								
a. Absorbers	24.9	25.6	26.1	24.2	26.3	26.9	29.0	35.9
b. Quenchers	24.9	25.3	22.0	25.2	25.2	24.3	27.2	33.1

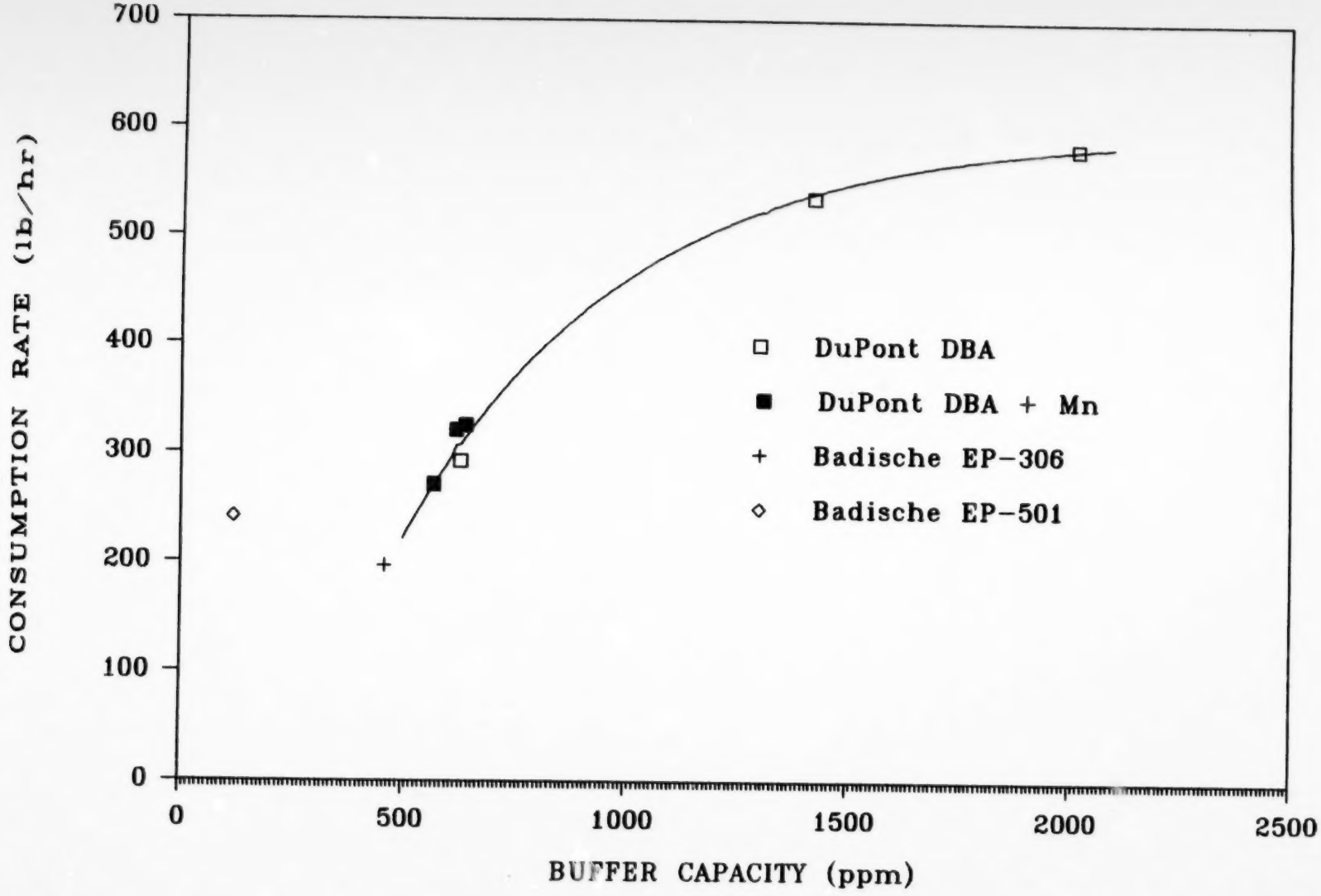

Figure 2 Acid Consumption vs Buffer Capacity

- Losses due to chemical degradation of the organic acids,

- Losses due to precipitation or co-precipitation of the organic acids with the calcium-sulfur salts produced in the system, and

- Losses due to organic acids present in process liquor entrained with the waste solids.

No attempt was made to differentiate or individually quantify these mechanisms.

Figure 2 illustrates several interesting results. First, it indicates that the DuPont DBA consumption rate increases at a less-than-linear rate over the range of buffer concentrations tested. In other words, a doubling of the DBA buffer capacity does not double the DBA consumption rate. This result is consistent with laboratory results for adipic acid which showed that consumption is proportional to acid concentration to the 0.55 power[4].

Another important result illustrated in Figure 2 is the effect of manganese on the DBA consumption rates. As shown, the DuPont tests with manganese do not differ appreciably from the comparable DuPont test without manganese. This is in contrast to results reported by other investigators. Previous studies have shown that manganese can reduce the chemical degradation part of adipic acid losses to near zero[4].

The reasons for the differences between the San Miguel manganese tests and previous tests are not clear. It may be that differences between pure adipic acid and the DBA acid mixture tested at San Miguel can account for the fact that manganese does not appear to have a significant effect on the DBA consumption rate. Another possibility is that the operating conditions during the manganese tests at San Miguel were outside the range in which manganese affects the organic acid degradation mechanism (e.g., the Mn concentrations were too low). Also, it may be that chemical degradation

represents only a small faction of the overall organic acid consumption rate at San Miguel. In any event, no benefit from manganese addition was observed at the conditions tested at San Miguel.

A third important result illustrated by Figure 2 is the relative consumption rates of the three organic acid additives tested at San Miguel. As shown, the two organic acid mixtures supplied by Badishe have significantly different consumption rates relative to each other. The additive EP-306, which is a mixture consisting primarily of mono-basic hydroxy caproic acid and dibasic adipic acid, shows a consumption rate which appears to be consistent with the consumption rates for the DuPont DBA mixture. However, EP-501 additive, which is a mixture consisting primarily of dibasic maleic acid, has a high consumption rate relative to the buffer capacity provided. Figure 2 shows that, although the buffer capacity for the EP-501 test is about 20 percent of that in the lowest buffer capacity DuPont additive test, the consumption rate is about 80 percent. Also, the consumption rate for the maleic acid test is actually higher than that for the EP-306 test while the buffer capacity is less than one-third of that provided by EP-306.

One final item regarding the organic acid consumption rates at San Miguel is the effect of inlet SO_2 concentration on consumption rates. Figure 3 presents data from San Miguel which indicate that the primary mechanism of organic acid loss may not be related to SO_2 absorption. This figure shows that, at equivalent buffer capacity, the SO_2 absorption rate does not appear to affect the organic acid consumption rate. The variation in SO_2 absorption rates shown in the figure is a result of a change in coal sulfur content (from 1.4 to 2.0). It should be noted in reviewing these results that the test with a high SO_2 absorption rate was conducted without manganese in the FGD system, while all the tests at low SO_2 absorption rates were coincidentally conducted with manganese. As a consequence, variables are confounded between the two groups of tests, and results should therefore be viewed with caution.

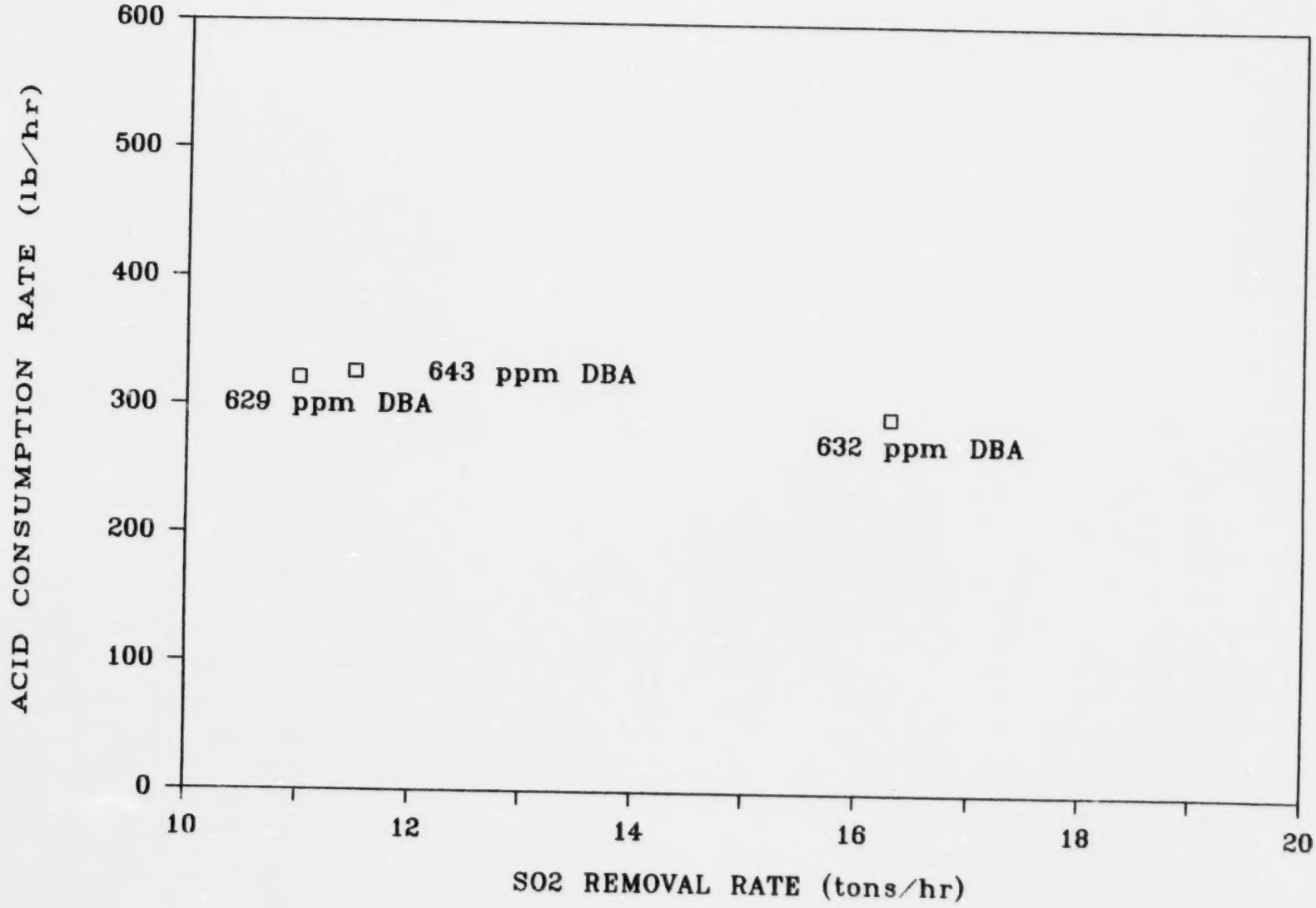

Figure 3 Acid Consumption vs SO_2 Removal

Limestone Utilization--

Limestone utilization is another factor which has a significant impact on the costs associated with organic acid addition. The major driving force for use of an organic acid additive at San Miguel was the potential for reducing limestone consumption by an amount which would more than offset the cost of the organic acid additive and the capital charges for an organic acid feed system. Some additional benefits from increased limestone utilization were also expected (e.g., reduced mist eliminator pluggage), but were not a significant factor in evaluating the costs of organic acid addition.

Table 4 summarizes the limestone utilization data collected during the test program at San Miguel. It includes data from baseline conditions (prior to organic acid addition) and each of the organic acid tests. Data are presented for both the quench and absorber loops, although the measurement of the quench solids represents overall limestone utilization since it is these solids which are ultimately discharged from the FGD system. As shown, the baseline utilization measured at San Miguel averaged 70 percent, while the organic acid addition tests had utilizations ranging from 83 percent to nearly 93 percent. The measured utilizations for the organic acid addition tests are based on samples collected from operating quench loops each day during the tests. This represents a substantial improvement which was accompanied by a reduction in process costs.

Operations Results

In addition to the quantitative results of the San Miguel organic acid tests, there were some qualitative observations made during the test program which may be of interest to utilities considering an organic acid additive to enhance the performance of a limestone FGD system. The discussion which follows summarizes those observations. Areas which are addressed include:

- pH meter maintenance and operation,
- Thickener operation,

TABLE 4. SAN MIGUEL TEST AVERAGES - OXIDATION, UTILIZATION, AND pH

Test ID	Absorber			Quencher		
	Oxidation	Utilization	pH	Oxidation	Utilization	pH
Baseline	20.0%	66.9%	6.16	18.9%	70.0%	6.09
DuPont 1	24.9%	85.9%	5.93	24.9%	85.0%	5.53
DuPont 2	25.6%	86.3%	5.87	25.3%	89.4%	5.49
DuPont 3	26.1%	74.0%	6.04	22.0%	91.3%	5.58
Badische	24.2%	65.7%	6.05	25.2%	82.7%	5.79
Maleic	26.3%	79.1%	6.16	25.2%	84.8%	5.84
Manganese 1	26.9%	79.3%	5.87	24.3%	87.8%	5.67
Manganese 2	29.0%	82.1%	5.85	27.2%	92.6%	5.64
Manganese 3	35.9%	82.3%	5.89	33.1%	88.4%	5.65

- Absorber and mist eliminator pluggage,
- Foaming in the recirculation tanks,
- FGD system water balance, and
- Corrosion of the DBA feed system.

A general discussion of each of these areas is provided along with comments relating to specific test conditions where appropriate.

pH Meter Maintenance and Operation--

Prior to the use of organic acid addition at San Miguel, pH was not a critical operating variable in the FGD system. Limestone addition was controlled by the pH feedback loop, but this was often overridden by the need to meet the 1.2 $lb/10^6$ Btu SO_2 emission limit. In such cases, limestone was added to the FGD system regardless of the pH. After the start of the organic acid addition tests, accurate pH readings became more important. It was necessary to control the quench loop pH at 5.4 to obtain the maximum cost benefit from the use of organic acid in the FGD system. Only by operating at this lower pH was it possible to reduce limestone stoichiometry and, therefore, reduce operating costs.

San Miguel has a regular pH meter maintenance program which includes calibration of the pH monitors once a shift and a check to ensure that there is adequate sample flow past the pH electrodes. Typical problems include pH meter drift, sample line pluggage, and breakage of pH electrodes. Of these problems, the last two are more serious. Plugged sample lines cause the pH meter to read high, while broken electrodes can cause erratic readings.

Because San Miguel had a pH meter maintenance program in place, system pH measurements were generally accurate. As a result, the pH meter maintenance schedule did not require any changes once organic acids were added to the FGD system. However, in many plants where the pH meters receive less

attention, an improved maintenance schedule will be needed to accrue the full cost savings associated with organic acid addition; and in some cases, an improved pH monitoring system may even be needed.

Thickener Operation--

One noticeable change in FGD system operation which occurred once organic acids were added to the San Miguel system was in the performance of the thickeners. Prior to the start of organic acid addition, the thickeners were often "cloudy," and they operated with only one clear sample port*. In addition, a flocculent was used to aid in setting of fine particles which were responsible for the "cloudy" appearance. Shortly after the start of organic acid addition to the system, thickener operation improved dramatically. Three clear ports in the thickener became common, and the addition rate of flocculent was reduced.

It is believed that the improvement in thickener operation is an indirect result of organic acid addition to the FGD system. The direct cause is the reduction in operating pHs which corresponds to a reduction in the amount of excess limestone in the system. At lower excess limestone levels in the FGD system, there is a reduced load on the thickener which would tend to improve operation. In addition, lowering the excess limestone would be expected to significantly reduce the number of very fine particles in the system. This should also help thickener operation.

The improvement in thickener operation discussed above was observed during all but one of the organic acid addition tests. The one exception occurred during the tests with the Badische DBA additive (EP-306). During these tests, foam from the quench tank was carried to the center well of the thickeners. The foaming action tended to cause the center well to overflow into the thickener and, as a consequence, the cloudiness in the thickeners

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*Sample taps are located in the side of each thickener at regular intervals. The number of "clear ports" is the number of taps from the top of the thickener which discharge clear liquor when sampled.

returned. This problem was one reason why the planned tests with EP-306 were cancelled.

Absorber and Mist Eliminator Pluggage--

A number of changes in FGD system operation were made possible once organic acid addition was initiated at San Miguel. As a result of those changes, several improvements in operation were observed. One of the most significant improvements was virtual elimination of scaling and plugging of the absorber trays and mist eliminators. This resulted in a significant reduction in the maintenance effort required to keep the FGD system running.

Prior to organic acid use, San Miguel routinely had a full time (40 hr/wk), 5 to 10 person crew cleaning absorber trays and mist eliminator passages in the spare module. However, after the start of the organic acid tests, module inspections revealed little or no scale in the absorbers or mist eliminator, and the cleaning effort was substantially reduced. The biggest problem which remained was pluggage of spray nozzles in the absorbers with deposits which formed and sloughed off the absorber and quencher walls.

The reduction of absorber tray and mist elimination plugging at San Miguel is attributed to the changes which were made possible by organic acid addition. Two of the most significant changes were:

- Operating at an absorber pH of 5.8 - At this pH there is less excess limestone in the recirculating slurry than during previous operation at a pH of 6.3 to 6.5.

- Operation at lower absorber gas velocities - Before organic acid addition, removal levels in the FGD system were just at the level required by regulations, and little flue gas bypass of the quench-absorber modules was possible. However, with organic acids, it became possible to bypass up to 10 percent of the flue gas around the scrubber modules. This lowered the gas velocity through the

modules and would be expected to reduce carryover to the mist eliminators.

Foaming in the Recirculation Tanks--

One problem which has been observed with the use of some organic acid mixtures in an FGD system is foaming. Foaming is caused by soap-like impurities in organic acid mixtures, and it can be a significant operating problem. During the tests at San Miguel, foaming was obseved under all conditions, but the severity of the foaming problem varied. During the tests with the DuPont supplied DBA, foaming was first observed at a buffer capacity of about 600 ppm. At this level of buffer capacity, the recirculating tank levels were reduced from 19 to 17 ft to provide 5 ft of freeboard in the tank and thereby prevent the tanks from foaming over. At higher levels of DuPont DBA, the foaming became more severe and, at a buffer capacity of 2,000 ppm, the recirculation tanks periodically overflowed. However, no other adverse effects of foaming were observed during this test.

Following the DuPont DBA tests, the FGD system was purged of organic acid by discontinuing addition of the DBA. Tests were then initiated using the organic acid material supplied by Badische Corporation (EP-306). During the first of three scheduled tests using this material, the foaming in the recirculation tanks was more severe at a buffer capacity of around 500 ppm than during the DuPont DBA tests at a 2,000 ppm buffer capacity. In response to this problem, San Miguel lowered the recirculatin tank levels an additional 2 ft (to 15 ft) which provided a total of 7 ft of freeboard. Even at this tank level, the Badische material caused the tanks to foam-over almost continuously.

FGD System Water Balance--

One minor effect of the changes made possible by organic acid addition was a reduction in the quantity of water consumed by the FGD system. At the lower limestone stoichiometries made possible with the use of organic acids, the quantity of waste generated by the FGD system was reduced by approximately 17 percent by weight. Assuming that the weight fraction of water in

the solids from the vacuum filter is relatively constant, a decrease in the total mass of solids results in a slight decrease in the total water consumed by the FGD system (i.e., evaporated water plus water associated with the waste solids). For San Miguel this change was estimated to be about 5 percent of the FGD system's total water requirement, and it did not have a significant impact on the system water balance. For other systems, the change in the system water balance may be more significant.

Corrosion of the DBA Feed Sytem--

San Miguel had a serious problem with corrosion of the organic acid feed system during the initial months of operation. Corrosion has occurred in all areas of the feed system including the tank heater (which failed after only 20 days of operation), lines, valves, and the tank itself. This problem has proved quite costly, increasing San Miguel's capital investment for the organic acid system by nearly 50 percent.

The DBA feed system at San Miguel was constructed of 304 SS. This type of steel has been used by Badische and DuPont for handling and storing DBA, and they have reported no problems. In addition, other FGD systems have used 304 SS successfully for handling DBA.

It appears, then, that the problems at San Miguel are site-specific. It is believed that San Miguel's corrosion problems are a result of mixing FGD process liquor and cooling tower blowdown (CTB) water with organic acids in the storage tank. An analysis of pipe samples from San Miguel's organic acid feed system has shown severe pit corrosion of the material. This corrosion was attributed to chloride attack of the metal under the acidic conditions (pH = 3.0) present in the DBA feed system. No chloride measurements were made of the DBA material itself, but an analysis of the FGD process liquor and CTB water showed 700 and 230 ppm chlorides, respectively.

Since the feed system corrosion became severe, San Miguel has replaced the stainless steel feed lines with PVC and has repaired the feed tank

twice. They have also avoided adding water to the organic acid feed. This has minimized the current problem, but obviously resulted in an additional capital expense. In summary, utilities considering the use of organic acids should benefit from this experience at San Miguel. The solution is to either avoid chloride addition to the organic acid feed or use a chloride resistent material such as fiberglass or PVC.

Additive Effects

Of the three organic acid additives tested, only the DuPont DBA was examined over a range of conditions. The Badische supplied acids (EP-306 and EP-501) were tested at only one level. The results of these tests permit two general comparisons. The DuPont material provided good performance over the range of conditions tested. However, foaming did begin to be a problem at a buffer capacity of 2,000 ppm. The Badische material EP-306 was tested only at a buffer capacity of about 500 ppm; but at this level, foaming was more severe than that observed for the DuPont material at 2,000 ppm. Because of the severe foaming, planned tests using EP-306 at higher buffer capacities were cancelled. The test with the Badische additive EP-501 did not exhibit any unusual foaming problems.

In terms of acid consumption rates, the DuPont material exhibited a less-than-linear increase in consumption as a function of buffer capacity. The Badische additive EP-306 appeared to have a consumption rate similar to the DuPont material. The additive EP-501 had a consumption rate which appeared to be significantly higher than that of either the DuPont material or the EP-306 additive. Comparisons of the consumption rates of these additives should be made with caution, however, because the tests with the Badische additives were outside the range of buffer capacities examined for the DuPont additive.

Economic Results

The overall measure of how organic acid additives affected operation of San Miguel's FGD system is the reduction in process operating costs which resulted. One method of evaluating the savings is to estimate the cumulative net present worth (CNPW) of the organic acid addition project. Table 5 summarizes the results of three, 15-year CNPW analyses which compare the three organic acid additives tested at San Miguel. Component contributions to the CNPW are also shown. The CNPW represents the estimated savings in mid-1984 dollars over a 15-year period of organic acid use in the San Miguel FGD system. As shown, the projected savings for the three additives range from nearly $3 million for the Badische EP-501 additive to $7.2 million for the DuPont DBA additive. This indicates that San Miguel will realize a very significant savings from the use of an organic acid additive. At current prices, the DuPont additive is the most attractive offering a 6 percent greater savings than the Badische EP-306 additive. Also, operation with the DuPont DBA provides additional benefits such as reduced absorber and mist eliminator pluggage and improved thickener operation. The full value of these benefits was not quantified in the CNPW analysis.

As shown in Table 5, the only cost items included in the CNPW analysis are the capital investments associated with the organic acid addition system and the continuing costs for the organic acid additive. The capital investment shown represents the initial investment in 1983, the costs incurred for repairs due to corrosion, and planned expenditures to improve the organic acid addition system in 1984. Since most of the capital costs have already been incurred, these will not be affected by the projected CNPW. However, a significant change in the delivered price of DBA or limestone could have a proportionally significant impact (either positive or negative) on the present worth of the organic acid addition project at San Miguel. This should be considered in using the San Miguel results to evaluate organic acid addition for other applications.

TABLE 5. FIFTEEN-YEAR CUMULATIVE NET PRESENT WORTH FOR THE THREE ADDITIVES TESTED AT SAN MIGUEL

Cost Component*	Acid/Buffer Capacity		
	DuPont DBA/ 600 ppm	EP-306/ 460 ppm	EP-501/ 125 ppm
Capital Investment	-215	-215	-215
Organic Acid	-6093	-3395	-5054
Limestone Reagent	6939	5949	4562
Limestone Grinding	1306	1120	859
Sludge Disposal	2523	2163	1659
Maintenance	752	---	---
Dewatering	353	303	232
Bypass Credit	1650	914	914
Total	7215	6839	2957

*All costs in mid-1984 dollars ($1000s)

Of the items in Table 5 which represent cost savings, most are related to the improvement in limestone utilization including:

- Limestone Reagent Cost Savings - This represents the improvement in limestone utilization which resulted from operation at lower quencher pHs. Baseline limestone consumption was estimated on 70 percent utilization.

- Limestone Grinding Cost Savings - The reduction in limestone grinding costs was calculated based on the estimated reduction in annual reagent consumption, costs for electricity, and the operating characteristics of the limestone grinding circuit. In addition, a reduction in the quantity of balls required to maintain the ball mill charge was estimated.

- Sludge Disposal Savings - Reducing the excess limestone produces a proportional reduction in the quantity of waste sludge generated by the FGD system. Savings associated with sludge disposal were estimated based on the limestone savings, operating characteristics of the dewatering area, and San Miguel's costs for waste disposal.

- Dewatering Cost Savings - A reduction in the mass of waste generated also reduces the load on the vacuum filtration system. This cost savings represents the reduction in power consumption which results from reduced operating time for the dewatering system.

There are two other items not related to limestone consumption which contribute to the projected cost savings at San Miguel:

- Reduced Absorber Maintenance - The use of DuPont DBA at a buffer capacity of 600 ppm allowed the San Miguel FGD system to operate at conditions which reduced the tendency of the absorbers

and mist eliminators to plug. San Miguel estimates that this has reduced the FGD system maintenance by at least 50 percent. The maintenance cost savings were not included for the two Badische additives because, when these additives were tested, the FGD system was operated at conditions which did result in plugging in the absorbers and mist eliminators.

- Bypass Credit - The use of organic acids in the San Miguel FGD system permitted varying amounts of flue gas to be bypassed around the modules. If more gas was bypassed, less reheat was required and a cost savings resulted. These savings were included in the economic analysis by estimating bypass rates for each organic acid addition test.

Bases for CNPW Analyses--

The results of the CNPW analyses are based on operating characteristics of the San Miguel power plant, FGD system performance data collected during the test program, actual costs for materials and services, San Miguel's escalation factors for projecting future materials costs, and the time value of money. As such, the CNPW results should represent savings San Miguel will realize in operation during the next 15 years.

The cost factors used in the CNPW analysis are shown in Table 6. The costs of organic acids and limestone represent prices FOB the San Miguel site. The electricity costs represent a bus-bar sales price, and the steam cost is based on San Miguel's fuel price and the thermal efficiency of the boiler. Other costs represent actual costs for materials or services. The cost escalation factors are also San Miguel's with 8.5 percent representing a petroleum related escalation factor, 7.5 percent the non-petroleum related factor, and 11.25 percent the interest rate. The use of the petroleum and non-petroleum factors for a given material or service was selected by Radian.

TABLE 6. UNIT COSTS AND ESCALATION FACTORS FOR PRESENT WORTH ANALYSIS

Cost Factor	Unit	Unit Cost ($)	Escalation Factor
Raw Materials			
DuPont DBA	lb	0.24	8.5
Badische EP-306	lb	0.19	8.5
Badische EP-501	lb	0.23	8.5
Limestone	ton	11.0	7.5
Utilities			
Electricity	kWh	0.04	8.5
Steam	10^6 Btu	1.33	8.5
Equipment Cost			
Original Installation	--	133,000	11.25
Retrofit Auto Control	--	50,000	
Corrosion Repairs	--	15,000	11.25
Labor Cost	hr	10.00	7.5
Interest	%	--	11.25
Miscellaneous			
Coal	10^6 Btu	1.00	8.5
Mill Balls	ton	500.00	7.5
Sludge Disposal	ton	2.00	7.5

Table 7 presents the calculational basis used in estimating annual costs and cost savings for specific unit operations. These numbers are, for the most part, based on actual operating levels at the San Miguel plant.

CONCLUSIONS

Several major conclusions can be drawn based on the results of the San Miguel test program:

- The use of organic acids in the San Miguel FGD system resulted in a significant improvement in process operation and a very substantial cost savings. The maximum estimated savings were $7.2 million in 1984 dollars over a 15-year period. These maximum savings were calculated for DuPont DBA additive added to attain an average buffer capacity of 600 ppm.

- Savings for the Badische additive, EP-306, were only slightly lower than for the DuPont material. However, the EP-306 additive was not "penalized" for the foaming problem, which made operation at a buffer capacity of more than 500 ppm very difficult. In fact, foaming was so severe that this test was terminated, and subsequent tests planned for higher buffer capacities were cancelled.

- Savings for the Badische additive, EP-501, were substantially less than for either the DuPont DBA or EP-306. This was due to a very high consumption rate of EP-501, relative to the buffer capacity provided.

- The key to realizing the savings associated with organic acid addition at San Miguel was a significant reduction in limestone stoichiometry in the FGD system. Limestone stoichiometry was reduced by reducing operating pHs in the quench and absorber recirculating loops.

TABLE 7. ASSUMPTIONS AND BASES FOR PRESENT WORTH ANALYSIS

Unit/Boiler	
Avg. Load	410 MW
Operating Time	7,200 hr/yr
Gross Heat Rate	10,500 Btu/kWh
Thermal Efficiency	75%
Flue Gas	
SO_2 To Scrubbers	7.08 lb/10^6 Btu
SO_2 Out Stack	1.10 lb/10^6 Btu
Stack Temp.	175°C
Heat In Bypass	2,582 Btu/10^3 scf
Reagent Preparation	
Mill Motor	134A, 6900V
Power Factor	0.88
Mill Throughput	46 tons/hr
Mill Ball Consumption	2,000 lb/day
Dewatering Plant	
Operating Power	651 kW
Solids Throughput	49 tons/hr
Filter Cake Solids	50%
Scrubber Maintenance Labor	
Annual Labor	12,600 man hrs
Limestone Utilization	
Baseline	70%
Baseline Tonnage	206,900 tons/yr
Fuel	
Heating Value (as received)	5,650 Btu/lb
Sulfur (as received)	2%

- Another benefit of operating at a lower absorber pH was a reduction in the pluggage of absorber trays and mist eliminator passages.

- Use of organic acids at San Miguel made some flue gas bypass possible. This contributed to the projected cost savings by reducing the requirement for reheat steam.

- Organic acid concentrations were not found to have a direct effect on the limestone utilization of the system. Only by allowing operation at lower pHs did the organic acids affect utilization.

- Tests with manganese added to the FGD system showed no effect on the organic acid consumption rate. This is in contrast to previous test results.

- Some of the test results from San Miguel indicate that organic acid consumption may, for a given FGD system, be independent of the SO_2 absorption rate.

In conclusion, substantial benefits resulted from the use of organic acids in the San Miguel FGD system. Significant cost savings will be realized by San Miguel due to changes in FGD operating conditions made possible by organic acid use. Also, improvement in FGD operation (such as reduced absorber and mist eliminator pluggage) was an indirect result of organic acid use at San Miguel. Because of these benefits, San Miguel plans to continue organic acid addition to the FGD system.

REFERENCES

[1] Hatfield, J.D., and Potts, J.M., "Removal of Sulfur Dioxide from Stack Gases by Scrubbing with Limestone Slurry: Use of Organic Acids," In: Proceedings of Second International Lime/Limestone Wet Scrubbing Symposium, APTD-1161, June 1972.

[2] Hargrove, O.W., et al., Full-Scale Utility FGD System Adipic Acid Demonstration Program: Volume I - Process Results, EPA-600/7-83-035a (NTIS PB83-238683), Radian Corporation, Austin, Texas, June 1983.

[3] Glover, R.L., A Cost Analysis of the Conversion of San Miguel Electric Cooperatives Limestone FGD System to Utilize Dibasic Acid: Draft Report, EPA Contract No. 68-02-3171, Task 56, Radian Corporation, Austin, Texas, 1982.

[4] Jarvis, J.B., et al., Effect of Trace Metals and Sulfite Oxidation on Adipic Acid Degradation in FGD Systems, EPA-600/7-82-067 (NTIS PB83-148379), Radian Corporation, Austin, Texas, December 1982.

CONVERSION FACTORS

To Obtain	Multiply	By
l	gal.	3.785
m^3/sec	10^3 ft^3/min	0.4719
l/s	gal./min	0.06308
kg	lb	0.4536
kg/l	lb/gal.	0.1198
metric tons	tons (short)	0.9072
g-mole	lb-mole	453.6
kW	hp	0.7457
Pascals	psia	6895
ng/J	lb/10^6 Btu	430
°C	°F-32	0.5556
J/kg	Btu/lb	2326
m	ft	0.3048
J	Btu	1055.1

TECHNICAL AND ECONOMIC EVALUATION OF ORGANIC ACID ADDITION TO THE SAN MIGUEL FGD SYSTEM

PART 2 - THE UTILITY'S PERSPECTIVE

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Organic-Acid-Enhanced Limestone FGD Process
San Antonio, Texas

July 19, 1984

TECHNICAL AND ECONOMIC EVALUATION OF ORGANIC ACID ADDITION TO THE SAN MIGUEL FGD SYSTEM

PART 2 - THE UTILITY'S PERSPECTIVE

ABSTRACT

San Miguel Electric Cooperative's FGD system started up in 1981 and immediately became an area of operating problems for the utility. After fighting major problems for over a year, San Miguel participated in an EPA sponsored study to explore the costs and benefits of an organic acid conversion to improve the performance of their limestone based flue gas desulfurization unit. Based on the results of the initial study, San Miguel agreed to participate with the EPA in a test program to gather actual data on the the conversion of a large commercial unit to organic acid enhanced operations. Several of the benefits include the following:

- Improved process operation and substantial cost savings were estimated to be $7.2 million over a 15-year period. Current operating experiences indicate that this figure is low. To ensure that these savings are real, San Miguel will continue to develop and refine its operating skills.

- Reduced limestone stoichiometry which is accomplished by reducing the operating pH and has resulted in better limestone utilization and slurry oxidation.

- Reduced scaling and pluggage in scrubber modules has increased efficiency with less maintenance.

- Reduced scrubber waste production requires less operation time for the system and less truck hauling time back to the mine.

TECHNICAL AND ECONOMIC EVALUATION OF
ORGANIC ACID ADDITION TO THE
SAN MIGUEL FGD SYSTEM

Part 2 - The Utility's Perspective

After San Miguel's start-up in 1981, it was immediately apparent that the flue gas desulfurization (FGD) system was going to be a problem area. In spite of close operation and maintenance supervision, many things occurred that increased the costs associated with this system as compared to our expectations. In addition, serious questions were raised about the availability and reliability of the system and our ability to operate within compliance.

For example:

(1) Limestone use averaged more than 200 tons per day above design. Stoichiometric ratios were in the range of 1.4 and 1.6, compared to the 1.2 design value.

(2) Severe scaling in the absorber trays and demister area required about 400 manhours per week for cleanout. This scaling also caused spray nozzle pluggage which, in turn, reduced scrubber performance to the extent that all four modules were required to be in service from time to time.

(3) The dewatering area operated three shifts per day with all three filters in service, yet was barely able to keep abreast of the incoming feed slurry. Design of this area was based on two shifts per day using two filters.

(4) Finally, increases in the use of auxiliary equipment resulted in reduction of the station net generation capacity.

After fighting these problems for over a year, it was agreed upon that there had to be a better way. That being the case, San Miguel decided to seriously consider the use of organic acid to improve the performance of their limestone based flue gas desulfurization unit. Prior to any expenditures, a study was requested by San Miguel, which was prepared by Radian Corporation for the EPA, to explore the costs and benefits of this conversion at San Miguel. The EPA has studied the use of organic acid addition to wet scrubbing systems since 1977. Based on an evaluation of the Radian study, published in a report dated October 1982, San Miguel agreed to participate with the EPA in a program to gather data on the technical and economic aspects of organic acid addition to a large commercial unit.

The findings of the Radian study presented reasons that San Miguel believed were credible enough to justify the investment. In general, the study found that:

(1) Reagent costs should remain about the same - less limestone consumption but additional costs for acid.

(2) Reliability and availability should increase due to less fouling, easier cleaning, less maintenance, etc.

(3) Bypass reheat would reduce fan horsepower and reheat steam consumption thus reducing fuel use or conversely allowing an equivalent increase in net generation.

FEED SYSTEM

The initial system cost estimate was $50,000.00 with an annual cost for acid additions of $250,000.00 at $0.24 per pound (dry basis) delivered price. Most of the commercially available organic acids contain some

percentage of nitric acid in addition to the various percentages of monobasic and dibasic acids. These organic acid solutions are delivered to the plant site in insulated tank trailers. The delivery temperature depends on the concentration. If the acid temperature is too low, crystallization will occur. The pH of the acid solution normally is below 2.0. These factors were considered for the materials of construction and the design of the system.

These solutions have been handled in series 300 stainless steels. Type 304L is preferred and was chosen as the basic construction material because of its durability over plastics and fiberglass.

The initial system consisted of:

- An insulated 30,000 gallon tank
- An immersion heater
- A variable speed gear pump
- An insulated and heat traced recirculating feed system. Each module had a flowmeter and needle valve at the absorber tank.
- A back pressure valve, pressure relief valve, by-pass valves, pump discharge valve, flush/drain valves, and flush hose connections.
- Temperature gages and tank level gage
- A water dilution and flushing system
- Plant air for truck unloading

San Miguel decided to use a 50 percent organic acid mixture because it can be stored at 104°F. This mixture can also be diluted to about 20 percent to allow for lower storage temperatures. If chlorides are present, corrosion of the stainless steel will occur. San Miguel experienced severe pit corrosion to the system after 3 weeks of operation due to chlorides in the dilution water. The tank bottom, tank immersion heater, and feed lines failed. The tank bottom was patched, the heater removed, and feed lines replaced.

The present system consists of the same tank with a new bottom and the same pump with a feed system running from the pump to the recycle water tank via thickener tank overflow. The feed rate of the 50 percent acid solution is monitored using rotameters with needle valves.

SYSTEM OPERATION

Receiving and Storage: The unloading valves are operated at the rear of the tank trailers. Quick connect hoses transport the material from truck to storage tank. Presently, San Miguel is unloading trailers using 20 psi air. No dilution water is being added to the organic acids.

Safety: Worker exposure to the acids has not created a problem. The item most often associated with using them is an unpleasant odor (valeric acid). Concentrated acid solutions may irritate eyes and skin. Hot concentrated solutions may cause thermal burns. Portable eyewash and water is available at the tank site in case of exposure or spillage. Burning or welding is prohibited in this area because of the possible release of nitrogen oxides above 277°F. Safety information is provided by each vendor.

Controls/Laboratory: The San Miguel tests have demonstrated that organic acid additions can be a cost-effective way to improve scrubber operations. Certain operating parameters have to be maintained to ensure that limestone utilization is optimized.

The parameters are:

(1)	Reagent limestone grind:	95% minus 325 Mesh
(2)	Reagent density:	30% Solids
(3)	Module density:	15% Solids
(4)	Module gas flow maximum:	525,000 ACFM
(5)	Acid concentration:	500-900 ppm
(6)	Quencher pH:	5.4
(7)	Absorber pH:	5.8

By maintaining the above operating conditions, a 10 percent by-pass credit can be achieved and still remain in compliance. The Chemical/Laboratory personnel monitor these conditions on a daily basis and coordinate with the Operations personnel. Organic acid concentration is measured using an automatic titrator, and the ppm data is used to adjust the feed rate. San Miguel is presently in the process of quantifying all of the benefits for operating with organic acids.

BENEFITS

Limestone Stoichiometry: Prior to the use of organic acids, limestone addition by pH feedback was frequently overridden to meet the 1.2 lb/MMBTU SO_2 emission standard. With the addition of these organic acids it became necessary to have better control over maintaining the proper pH to ensure that maximum benefits could be achieved. Operating at the tower pHs of 5.4 (quencher) and 5.8 (absorber) makes it possible to reduce limestone stoichiometry. Before organic acid, it was not infrequent to have

stoichiometry above 1.4; now it is 1.2 or lower. Maintaining these lower stoichiometries equals savings in limestone up to $90,000.00 a month. Process costs savings are also realized.

Reheat: As soon as the additions of organic acid were made, it became possible to by-pass flue gas, thus reducing reheat steam requirements. This reduction in steam consumption amounts to 1 ton/hr of fuel saved or $100,000.00 a year.

Presently San Miguel is operating at approximately 10 percent by-pass. This setting allows for a savings in I.D. fan horsepower which equate to 1.4 MW/hr or an additional savings of over $350,000.00 a year.

Waste Disposal: By reducing the excess limestone and obtaining better utilization, less waste sludge was generated. This amounted to approximately 17 percent.

Prior to organic acid, it was difficult to maintain one clear port in the thickener tanks. Ports are sample taps located in the side of each thickener which are arranged in regular intervals from top to bottom. A flocculant is metered into each thickener tank to aid in the settling of finer particles. With organic acids it is possible to have three clear ports with reduced flocculant additions.

The thickener solids separation is much faster, decreasing handling time. Scrubber waste operation has been reduced from 20 hours to 12 hours. Instead of running three shifts a day, it is now possible to process waste products in two shifts a day.

Because of the more complete oxidation of solids (from 20 percent to 30 percent) and better dewatering properties of these solids, less fly ash has to be added to the waste sludge. Radian Corporation has projected a $212,000.00 savings per year for this part of the FGD system.

Scrubber Cleaning: As mentioned in the beginning of this paper, San Miguel experienced severe scaling problems in the demisters and on the absorber trays, along with nozzle pluggage. These required over 400 manhours per week for maintenance and cleaning. At times it was necessary to operate all four modules to stay within compliance, further compounding these problems.

After organic acid additions began, five men can clean a module in two days, amounting to a 320 hour reduction in scrubber maintenance. This has come about because of the reduced scale buildup and reduced nozzle pluggage.

OPERATIONS BENEFITS

Absorber Trays: Modules stay cleaner because less gas is processed and less excess limestone is present at lower pH. Less alkali is available to form soft calcium sulfite and hard calcium sulfate deposits. This has reduced the scaling and plugging of absorber trays, and plugging of nozzles.

Demisters: For these same reasons there is considerably less pluggage of the chevrons.

Nozzles: Pluggage of nozzles has been greatly reduced, but still remains a problem. Scale falls from the walls and is pumped into the nozzles.

Thickeners/Dewatering: Organic acid additions have resulted in maintaining a lower sludge blanket in the thickeners, which is evident through observation of more clear ports. Oxidation of sludge waste products has increased from 20 percent to 30 percent. Underflow percent solids has increased from 30 percent to 35 percent. As a result, improved vacuum filtering capabilities now exist. Filter cake solids have increased from 50 percent to 60 percent.

CONCLUSION

The benefits which have been realized from the testing program have expanded the technical and economic data base for FGD organic acid addition.

Presently, San Miguel is in the process of gathering more data and quantifying these results.

Several of these benefits have been discussed above, but in summary they are:

(1) Improved process operation and substantial cost savings were estimated by Radian Corporation to be 7.2 million dollars over a 15 year period. Current operating experiences indicate that this figure is low. To ensure that these savings are real, San Miguel will continue to develop and refine its operating skills.

(2) Reduced limestone stoichiometry, accomplished by reducing operating pH, has resulted in better limestone utilization and slurry oxidation.

(3) Reduced scaling and pluggage in scrubber modules has increased efficiency with less maintenance.

(4) Reduced scrubber waste production requires less operation time for the system and less truck hauling time back to the mine.

San Miguel still has to quantify some of the additional benefits, but it is obvious at this time that organic acid additions can alleviate many problems associated with limestone based flue gas desulfurization systems.

RESULTS OF THE FIRST TWO YEARS OF COMMERCIAL OPERATION OF AN ORGANIC-ACID-ENHANCED FGD SYSTEM

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ABSTRACT

The U. S. EPA has sponsored research to develop organic-acid-enhanced flue gas desulfurization (FGD) technology for existing and new coal burning facilities. A 1981 EPA-sponsored demonstration program at Springfield City Utilities' Southwest Power Plant (SWPP), near Springfield, MO, showed that adipic acid and dibasic acid (DBA) greatly enhanced FGD performance. SWPP has continued to use DBA to comply with the 1971 SO_2 emissions standard under which they are regulated. Thus, SWPP became the first commercial-scale system to use an organic additive to enhance SO_2 removal. This paper documents the first 2 years (1982 and 1983) of commerical operation of the DBA system at SWPP. During 1982 and 1983, SWPP averaged an SO_2 emission rate of less than 1.0 lb $SO_2/10^6$ Btu (430 ng/J). Conversely, in 1980 (prior to DBA addition), SWPP averaged about 5 lb $SO_2/10^6$ Btu (2,170 ng/J). FGD system reliability has also greatly improved, averaging 97.9 percent in 1982 and 98.7 percent in 1983, compared to 45 percent in 1980. Overall, DBA has increased the flexibility of the SWPP system and, most importantly, allowed SWPP to operate in compliance.

INTRODUCTION

EPA's Industrial Environmental Research Laboratory at Research Triangle Park (IERL-RTP) has an ongoing research and development program to improve the performance of lime/limestone flue gas desulfurization (FGD) systems. One of the most promising aspects of that effort has been the use of organic acids as buffering agents to increase soluble alkalinity in limestone FGD systems. Adipic acid was identified as a suitable additive because of physical properties, availability, and cost. After extensive laboratory-, pilot-, and prototype-scale studies on adipic-acid-enhanced limestone systems, EPA, in 1981, sponsored a full-scale demonstration at SWPP.

Nine months of testing was conducted at SWPP. Four months were spent at the beginning of the program investigating adipic acid addition in a forced oxidation system. Better than 90 percent SO_2 removal was achieved in

a 49-day run, the longest continuous operating period at SWPP up to that time. Next, adipic acid addition achieved 85 - 95 percent SO_2 removal in a natural oxidation system over a period of 3½ months.

Finally, a mixture of organic dibasic acids (DBA) was tested over a 6-week period following the adipic acid testing. DBA, a by-product of adipic acid manufacturing, contains adipic, glutaric, and succinic acids, and is less expensive than adipic acid. Testing at SWPP showed DBA to be very effective in enhancing FGD system performance, resulting in a significant decrease in SO_2 emissions.

An analysis was then performed to determine the most cost effective option which would enable SWPP to comply with the 1971 SO_2 emissions standard. Options considered included increasing the slurry rate to the scrubbers, converting the scrubbers to spray towers, adipic acid enhancement, and DBA addition. Lower costs associated with the DBA option, as well as its demonstrated effectiveness, led SWPP to select DBA for long-term use. Thus, SWPP became the first commercial-scale facility to utilize DBA in an FGD system application. This paper documents commercial operation of SWPP's DBA system, between January 1982 and December 1983.

PROGRAM OBJECTIVES

Since operation at SWPP represents the first commercial scale use of an organic-acid-enhanced FGD system, IERL-RTP personnel decided it would be useful to the industry to document system performance following the demonstration program. Consequently, in May 1982, Radian Corporation was contracted to monitor SO_2 emissions and the overall process conditions at SWPP. The primary objectives of this program were to evaluate the performance of the DBA-enhanced system and to note any long-term effects on the process. A brief description of the SWPP system and a summary of the monitoring program results are presented here.

FGD SYSTEM DESCRIPTION

The SWPP FGD system consists of two parallel tray-tower modules, each sized to handle 60 percent of the design flow. The coal typically contains 3.5 - 4.0 percent sulfur, corresponding to inlet flue gas concentrations of 2,000 - 2,600 ppm SO_2. Limestone preparation, sludge dewatering, and sludge disposal systems are common to the two modules.

Figure 1 shows the overall flow scheme for the FGD system. The flue gas enters the presaturator, where it is cooled by clarified liquor (supernatant) from the thickener. The gas then passes through gas/liquid contact trays, the mist eliminator, and out of the stack. No flue gas reheat is used.

Slurry is circulated from the reaction tank below the scrubber through spray nozzles above the top level of trays. Makeup limestone is added from the limestone storage tank to the reaction tank to maintain the pH set point. Solids are first settled in the thickener, and the thickener underflow slurry is dewatered on a vacuum filter belt. The resulting sludge is mixed with fly ash in the pug mill and trucked to an on-site landfill. Supernatant (thickener overflow) from the dewatering operation is returned to the FGD modules as presaturator and mist eliminator wash water. Additional supernatant water is utilized in the limestone preparation area.

In December 1981, a temporary DBA feed system was installed to add DBA solution on a continuous basis. DBA was delivered on site in 6,000 gal. (22,710 liter) tank trucks. Hot water (approximately 140°F, 60°C) was circulated through a steam jacket surrounding the tank to inhibit DBA crystallization.

Initially, DBA was pumped from the tank truck to the ball mill sump. The feed rate was set manually. After the DBA was mixed with the freshly ground limestone slurry from the ball mills, the mixture was pumped to the

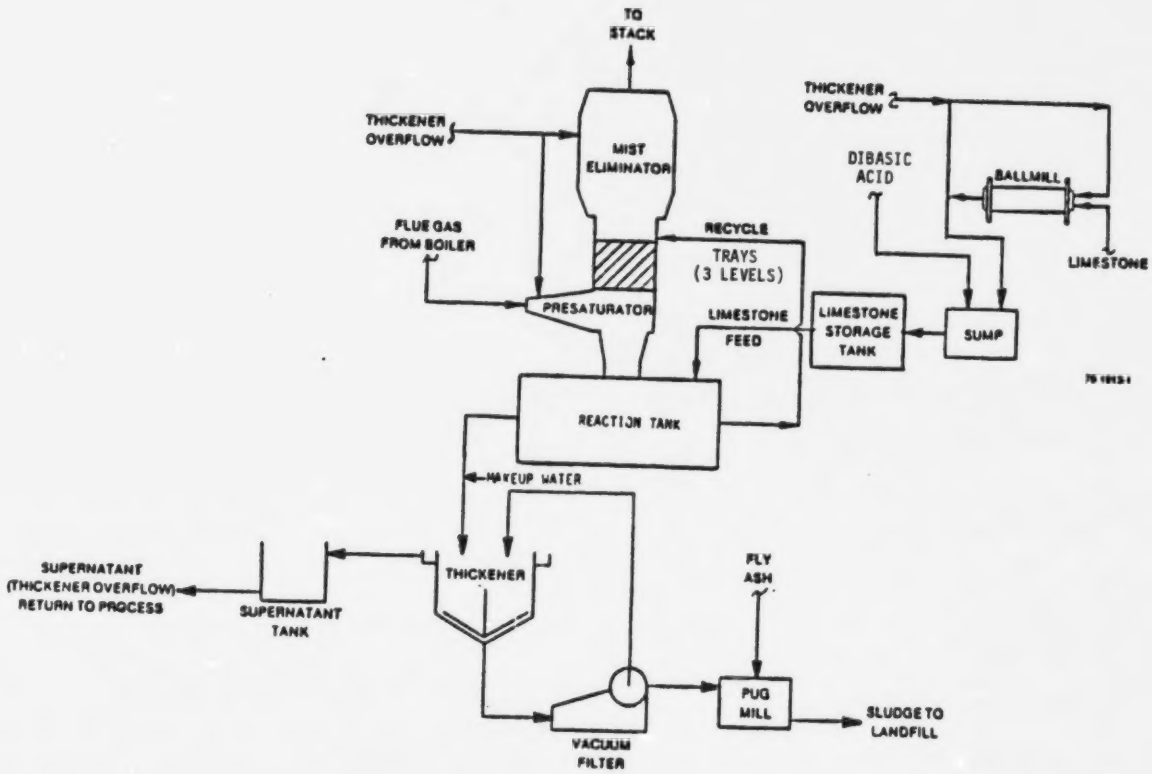

Figure 1. SWPP FGD Flow Scheme, 1981.

limestone storage tank. Since the limestone handling equipment was common to both scrubber modules, both scrubbers operated at essentially the same DBA concentration. Later (in September 1982), the DBA feed system was modified to allow DBA to be fed separately into the reaction tank of either scrubber.

In May 1983, SWPP began operating with a permanent DBA feed system, including an 18,000 gal. (68,000 liter) stainless steel storage tank, an immersion heater, two 30-gpm (114 l/min) DBA circulation pumps, feed lines to the individual modules, and a source of heated flush water (see Figures 2 and 3). The cost of the permanent system was $275,000. To date, SWPP has reported no major problems with the permanent DBA feed system.

PROGRAM RESULTS

Analysis of the SWPP FGD system operation in 1982 and 1983 shows that the DBA-enhanced system performed well. SO_2 emissions averaged less than 1.0 lb $SO_2/10^6$ Btu (430 ng/J) in 1982 and 1983, compared to about 5 lb $SO_2/10^6$ Btu (2,170 ng/J) in 1980 (see Figure 4). The average SO_2 removed also increased in 1982 and 1983 as compared to 1980 (see Figure 5). FGD system reliability averaged over 97 percent in 1982 and almost 99 percent in 1983, compared to 45 percent in 1980, before the adipic acid demonstration program (see Table 1).

The flexibility of the DBA-enhanced system was well demonstrated in 1982 as SWPP personnel adjusted conditions to minimize scaling in the mist eliminators before installing limestone classifiers. The pH set point was reduced, from the 5.4 used during the DBA demonstration, to 5.0. The reduced pH resulted in increased limestone utilization which helped maintain mist eliminator reliability. Higher DBA concentrations were required with the lower pH setting, which resulted in a higher DBA consumption rate; i.e., 30 lb DBA/ton SO_2 removed (15 g DBA/kg SO_2) compared to 15 lb/ton (7.5 g/kg) in the demonstration program. However, system reliability was maintained at

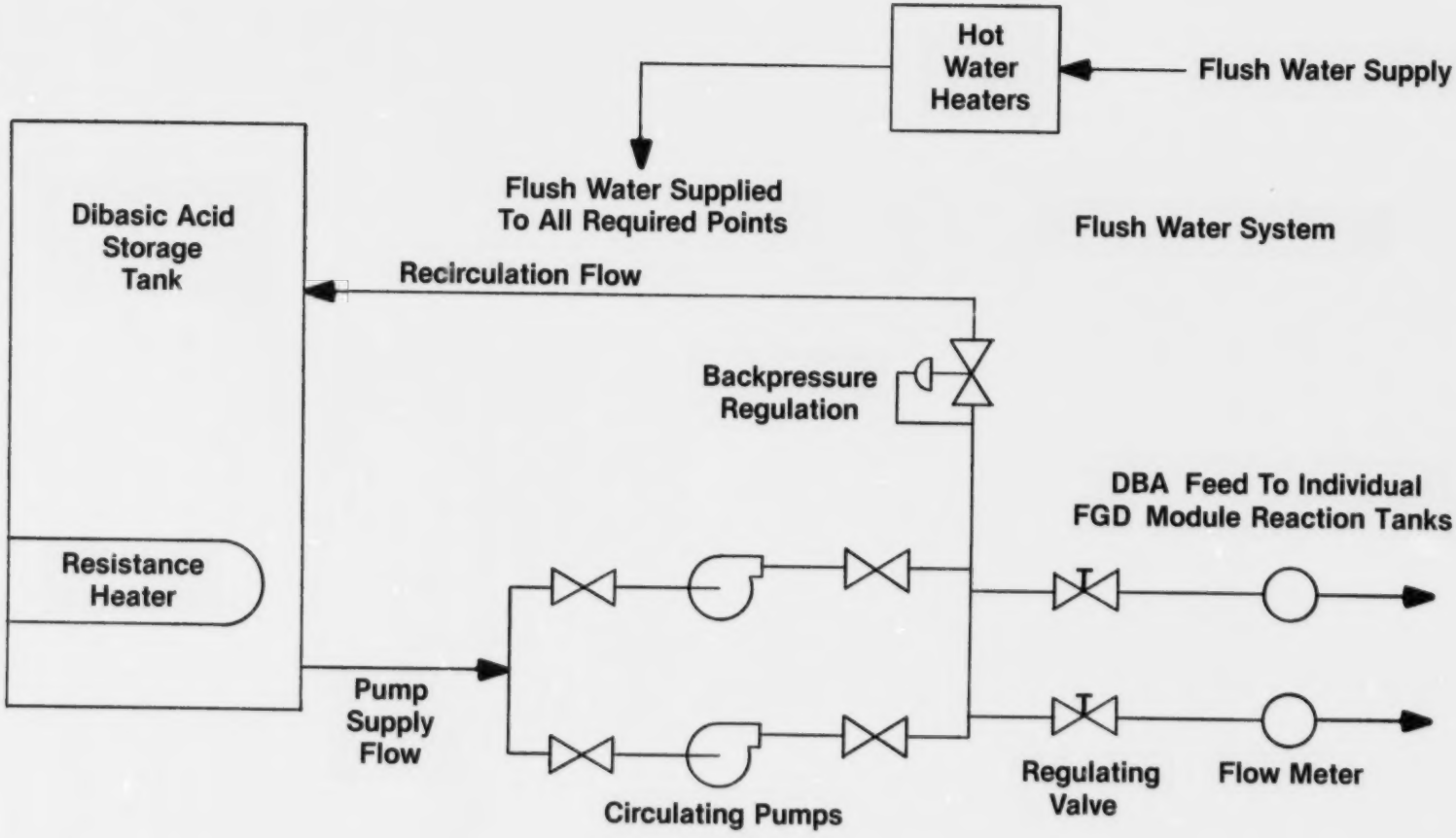

Figure 2. DBA Flow Schematic for Permanent System

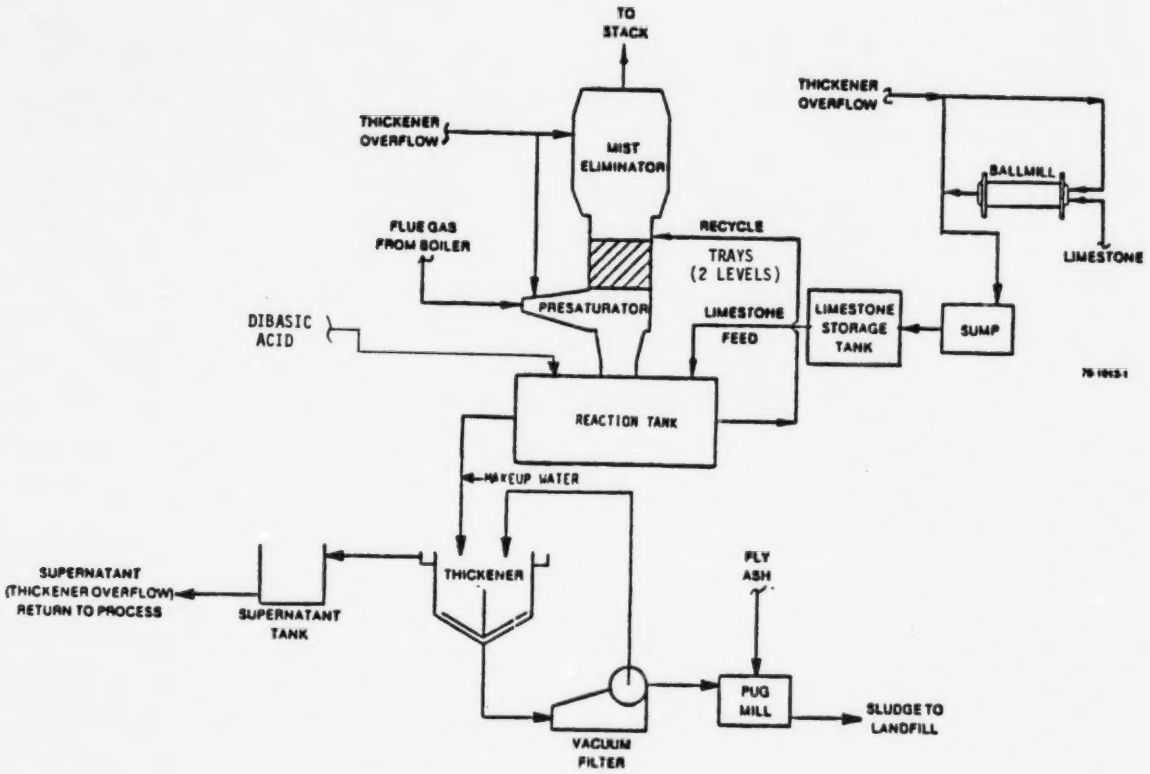

Figure 3. SWPP FGD Flow Scheme, 1984.

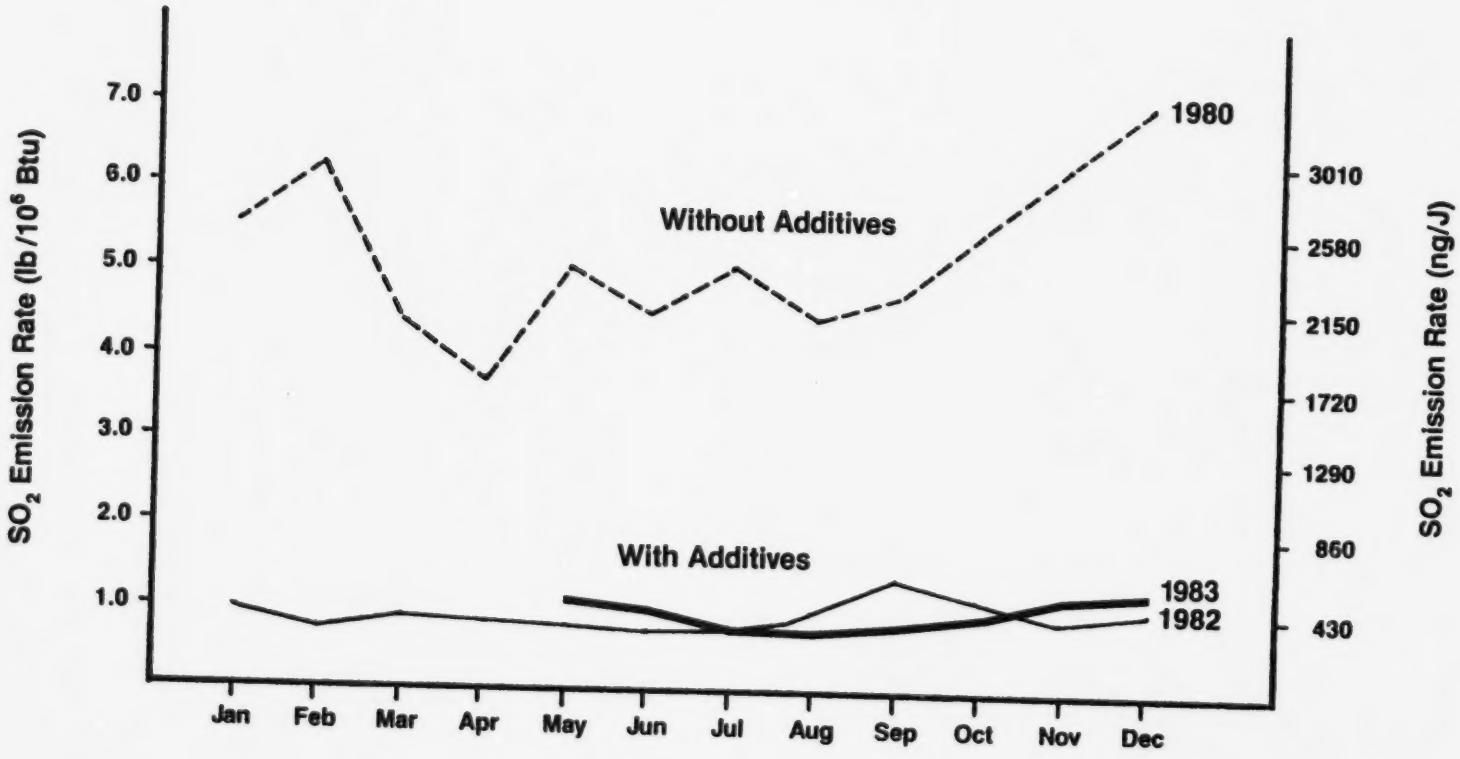

Figure 4. Comparison of SWPP Monthly Average Emission Rates With and Without Organic Acid Additives.

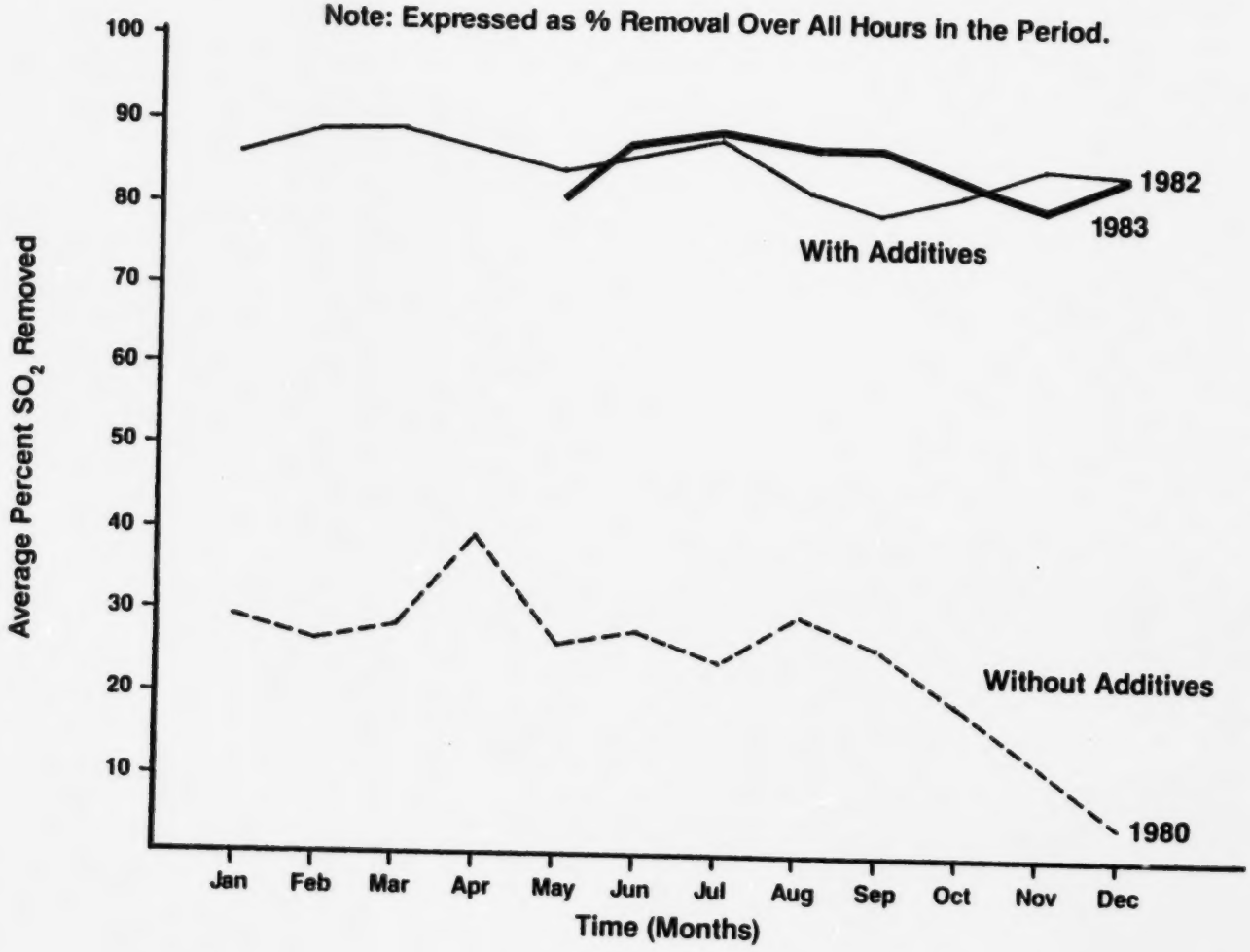

Figure 5. Comparison of SWPP SO_2 Removal Efficiencies With and Without Organic Acid Additives

TABLE 1. SWPP FGD SYSTEM RELIABILITY IN 1982 and 1983

Month	Reliability[1] 1982	1983
January	98.6	-outage-
February	99.8	-outage-
March	98.5	-outage-
April	-outage-	-outage-
May	96.4	96.1
June	97.8	99.9
July	98.0	99.1
August	96.7	98.5
September	99.9	99.8
October	96.0	99.7
November	97.6	99.6
December[2]	92.8	96.7
Average	97.5	98.7

[1] Reliability design = $\frac{\text{Hours FGD system operated}}{\text{Hours called upon to operate}} \times 100$

[2] Scheduled outage started 12/3/82. Down a short time due to lightning.

Source: SWPP Quarterly Status Reports submitted to U. S. EPA Region VII.

a very high level in spite of a poor limestone grind. The installation of limestone classifiers in late 1982 permits operation at higher pH with good limestone utilization. 1983 results reported by SWPP show that DBA consumption was reduced to about 25 lb per ton SO_2 (12.5 g/kg SO_2) using the new classifier system. Further reduction in DBA consumption may be possible.

CONCLUSIONS

SWPP's FGD system continues to perform very well with respect to both reliability and SO_2 reductions. DBA has increased the versatility of the FGD system as shown by the successful operation at pH 5.0. SO_2 emissions were well within the 1.2 lb $SO_2/10^6$ Btu (520 ng/J) limit, and the reliability of the FGD system throughout 1982 and 1983 was excellent. The following conclusions can be drawn from this program:

- Use of DBA allows SWPP to remain in compliance with the SO_2 regulation. An average SO_2 emission of less than 1.0 lb $SO_2/10^6$ Btu (430 ng/J) was measured in 1982 and 1983 with DBA addition, compared to 5 lb $SO_2/10^6$ Btu (2,170 ng/J) in 1980, before the acid addition demonstration program.

- DBA greatly increases the flexibility of the FGD system, allowing successful operation over a wide pH range. Operation at a reduced pH, in conjunction with better utilization of fresh water to the FGD system, allowed SWPP to achieve an overall system reliability of 97.5 percent in 1982, and 98.7 percent in 1983.

- Limestone classifications and grinding circuit performance are important to overall system performance. In 1982, DBA consumption at SWPP was about 30 lb/ton SO_2 (15 g/kg SO_2), due primarily to operating at a lower pH and a decrease in recycle pump performance. A limestone classification system was installed in late 1982, enabling the system to operate at a pH of 5.6 - 5.8 with good limestone utilization. Higher pH operation reduced the DBA consumption rate to about 25 lb/ton SO_2 (12.5 g/kg SO_2) in 1983. Further reductions in DBA consumption may be possible.

NOTE: This paper is based on the Project Report, "Results of the First 2 Years of Commercial Operation of an Organic-Acid-Enhanced FGD System," EPA-600/7-84-065, June 1984. This work was sponsored by EPA under Contract 68-02-3171, Work Assignment No. 53 with Radian Corporation.

ECONOMICS OF RETROFITTING BIG RIVERS ELECTRIC CORPORATION'S LIME BASED FGD SYSTEM TO ORGANIC ACID ENHANCED LIMESTONE OPERATIONS

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ECONOMICS OF RETROFITTING BIG RIVERS ELECTRIC CORPORATION'S LIME BASED FGD SYSTEM TO ORGANIC-ACID-ENHANCED LIMESTONE OPERATIONS

ABSTRACT

In 1982-83, Peabody Process Systems, Inc. (PPSI) conducted pilot plant tests at the R. D. Green Station of Big Rivers Electric Corporation (BREC). PPSI's final report of the pilot testing included comparisons of the operating costs of a lime based, full-sized absorber, to a retrofit limestone system enhanced with dibasic acid (DBA) or adipic acid. The site specific changes required for BREC to convert their existing lime FGD system to a limestone system enhanced by DBA or adipic acid, and the costs of making such a change are described in this paper. Results of this analysis indicated that an annual cost savings of $2.6 million could be achieved by converting the existing lime system to an adipic-acid-enhanced limestone system, and an annual savings of $3.1 million could be achieved by converting to a DBA-enhanced system.

1.0 INTRODUCTION

In 1982-83, Peabody Process Systems, Inc. (PPSI) conducted pilot plant tests at the R. D. Green Station of Big Rivers Electric Corporation (BREC). PPSI's final report of the pilot testing included comparisons of the operating costs of a lime based, full-sized absorber, to a retrofit limestone system enhanced with dibasic acid (DBA) or adipic acid. The economics did not include the cost of new equipment required to make the conversion. A second project was funded by the Environmental Protection Agency (EPA) in order to define the capital requirements necessary for the retrofit and thereby complete the economic comparison.

The existing R. D. Green Station has two 240 MW boilers that use an American Air Filter (AAF) Flue Gas Desulfurization (FGD) system employing a spray tower absorber and dolomitic lime as a reagent. Dolomitic lime enhances SO_2 removal efficiency through the accumulation of dissolved alkalinity ($MgSO_3$), which allows a relatively low liquid-to-gas ratio (L/G) to be used for the system. Adipic acid and DBA enhance the performance of limestone scrubbing systems such that a mixture of limestone and DBA or adipic acid may be substituted for lime with minor changes to an existing FGD system.

This paper addresses the site specific changes required for BREC to convert their existing lime FGD system to a limestone system enhanced by DBA or adipic acid, and the costs of making such a change. Results of this analysis indicated that an annual cost savings of $2.6 million could be achieved by converting the existing lime system to an adipic acid enhanced limestone system, and an annual savings of $3.1 million could be achieved by converting to a DBA enhanced system.

2.0 BIG RIVERS, R. D. GREEN STATION - SYSTEM DESCRIPTION

The Big Rivers Electric Corporation's Green Units No. 1 and 2 each consists of a coal-fired steam generator with an electrostatic precipitator (ESP) followed by a wet sulfur dioxide (SO_2) absorber system. The steam generator is a Babcock and Wilcox balanced draft, pulverized-coal-fired, Carolina-type Radiant Boiler. The maximum continuous steam flow rate is 1,930,000 lb/hr[(1)] at 1975 psig superheater outlet pressure with 1,005° F main steam temperature. Two Ljungstrom Regenerative Tri-Sector air heaters transfer heat from the exit flue gas to the entering combustion air.

The flue gas exiting the air heaters is cleaned of particulate in an AAF ESP designed to remove 99.64 percent of the maximum expected load of 75,000 lb/hr of flyash. To achieve this, AAF installed a two-unit ESP: each unit has 48, 12-in. wide gas passages and collection plates 42 ft high. The gas passages have an effective depth of 55 ft, 5 in. (consisting of seven 7 ft 11 in. sections). Each ESP unit has two parallel electrical sections, which are seven fields deep.

AAF has also designed and manufactured the SO_2 absorption system. This system uses an aqueous slurry of dolomitic lime to react with the SO_2 and produce a calcium sulfite and sulfate slurry. The system consists of two spray tower reaction modules for each boiler, along with their associated pumps, valves, and controls. The system is designed to handle all of the flue gas from the steam generators and remove 90 percent of the SO_2. The absorbers are equipped with a bypass duct and damper arrangement which can be used to bypass flue gas in an emergency. Each module was designed to handle 60 percent of the rated full-load flue gas flow. Flue gas exiting

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(1) Environmental Protection Agency policy is to express all measurements in Agency documents in metric units. In this paper, however, English units are used to avoid lack of clarity. Conversion factors from English to metric units are given in Table 9 located at the end of this paper.

the absorbers is reheated by heating ambient air in a steam-to-air heat exchanger and forcing the hot air into the duct connecting the spray towers and the stack with a separate reheat fan.

The lime slurry is prepared in a lime slaker and pumped to the absorber recycle tank in response to decreases in pH. The pH is normally maintained at 7.0. The suspended solids are maintained at about 2.0 percent solids by the addition of a supernate from the thickeners and makeup water to the system. The waste slurry from the recycle tank, typically containing 2 to 3 percent solids, is fed to the four thickeners for primary dewatering. The supernate is returned to the scrubber system. The thickener underflow contains 20 percent solids and is fed to two of the three rotary drum vacuum filters for secondary dewatering. The vacuum filters discharge a 48 percent solids cake. This cake is blended with flyash and lime in a pugmill blender to produce a final product containing 60 to 65 percent solids. Over time, chemical reactions of the flyash and lime in the mixture produce a stable rock-like material suitable for a landfill. The key design parameters for the air quality control systems (AQCS) are summarized in Table 1.

Due to high water content in the filter cake, nearly all flyash produced by the two Green Units must be used to produce a final mixture with sufficient solids content (60 to 65 percent) for easy handling and transporting to the landfill. This solids content also ensures stable landfill material with relatively high strength and low permeability.

3.0 CONVERSION TO ORGANIC ACID ENHANCED OPERATIONS

3.1 Prediction of Full-Scale SO_2 Removal Efficiencies with Limestone/Organic Acid

The BREC pilot plant test results could not be used to predict SO_2 removal efficiencies for the full-scale spray tower since wall effects reduced the spray zone mass transfer effectiveness for the pilot plant

TABLE 1. AQCS DESIGN PARAMETERS BREC'S GREEN STATION UNIT NO. 2[a]

ESP	
Design collection efficiency	99.64%
Design pressure drop	0.75 in. W.G.
Design flow rate	1,000,000 acfm @ 300°F
Average face velocity	3.95 ft/sec
Treatment time	12.02 sec (6 fields)
Expected outlet dust loading	0.03103 gr/acf
Number of units	2
Number of gas passages	48
Number of fields	7
Plate height	42 ft
Plate width	7 ft, 11 in.
Plate spacing	12 in.
Total collection area (7 fields)	468,156 ft^2
Collection area with 6 fields (guarantee condition)	401,277 ft^2
Design SCA (6 fields)	400 ft^2/1,000 acfm
Design migration velocity (6 fields)	7.12 cm/sec
Guaranteed average power consumption (6 fields)	860 kW
FGD SYSTEM	
Design flow rate	1,000,000 acfm @ 300°F
Design SO_2 removal efficiency	90%
Number of modules	2
Tower diameter	34 ft

(continued)

[a]Unit No. 1 is a mirror image of Unit No. 2 and has the same AQCS design parameters.

TABLE 1. (Continued) AQCS DESIGN PARAMETERS BREC'S GREEN STATION UNIT NO. 2

FGD SYSTEM (Continued)	
Spray zone height	26 ft
Number of spray levels	6
Sulfur in coal	4.0%
SO_2 to scrubbers	22,167 lb/hr
SO_2 absorbed	19,950 lb/hr
Design flow rate per module, saturated	446,200 acfm
Saturation temperature	128°F
Design velocity	8.2 ft/sec
Design recirculating slurry (per module)	21,275 gpm
Modified recirculating slurry (Module A)	14,900 gpm
Modified recirculating slurry (Module B)	16,500 gpm
Design total lime feed	19,133 lb/hr
Recycle tank volume	183,600 gal.
Number of thickeners	4
Thickener diameter	125 ft
Thickener volume	898,650 gal.
Design thickener capacity	37 ft^2/tpd dry solids
Design underflow suspended solids	30% (actual 20%)
Design underflow flow rate	160 gpm
Filter feedrate	291 gpm/filter
Design filter capacity	70 lb/hr/ft^2
Filter drum diameter	12 ft
Filter drum length	20 ft
Number of filters required per unit	2 + 1 spare
Cake moisture	48%
Flyash feed rate to pug mill	70,400 lb/hr
Lime feed rate to pug mill	4,000 lb/hr

(Continued)

TABLE 1. (Continued) AQCS DESIGN PARAMETERS BREC'S GREEN STATION UNIT NO. 2

FGD SYSTEM (Continued)	
Lime stoichiometry	1.02 moles lime/mole SO_2 removed
Final composition flyash/ filter cake weight ratio	0.6/1
Lime in final mixture	2%
Final solids content, actual	60 - 65%
Final solids content, optimum	70 - 75%

operations. An attempt was made to compare the existing BREC spray tower performance to the performance of the 10 MW Shawnee spray tower. Typically, larger diameter spray towers are more efficient than the 10 MW Shawnee spray tower; consequently, use of the 10 MW data for predicting full-scale performance should be conservative.

However, a direct comparison of the operation of the full-scale BREC spray tower to the Shawnee operations could not be made since much of the required process data (Mg^{++} concentration, pH, L/G, inlet SO_2 concentration, and temperature) were not available. For this reason, semi-empirical equations developed by Bechtel[1] for limestone scrubbing with adipic acid were used to estimate full-scale liquid pumping requirements for achieving 90 percent SO_2 removal as a function of inlet SO_2 concentration.

Adipic acid and liquid pumping requirements for the retrofit were calculated for several inlet SO_2 conditions. Increased L/G rates would be obtained by increasing the rpm of the existing recycle pumps to the capacity of the existing motors. DBA requirements would be the same as adipic acid when expressed as parts per million of equivalent adipic acid.

Table 2 shows the minimum L/G required to obtain 90 percent removal for different inlet SO_2 concentrations. As indicated, even at the highest sulfur coal, there is sufficient pumping capacity for meeting compliance when 3,000 ppm of adipic acid is used.

3.2 Basis for Retrofit Equipment Design

The ultimate coal analysis typical of coal burned at BREC (Table 3) was used as a basis for performing combustion calculations. These combustion calculations assumed an excess air rate of 50 percent and a coal feed rate of 277,000 lb/hr which were consistent with operating conditions measured by Engineering Testing Services (ETS) in a series of tests to evaluate the performance of BREC's ESP in January 1982. Combustion calculation results

TABLE 2. CALCULATED LIQUID TO GAS RATIOS REQUIRED TO OBTAIN 90 PERCENT REMOVAL[a]

Adipic Acid Concentration, ppm	L/G Ratios for Two Inlet SO_2 Concentrations, ppm[b]	
	2,500	4,000
0	103 gal./10^3 acfm	146 gal./10^3 acfm
1,000	88	125
2,000	51	72
3,000	29	42

[a]Calculated using semi-empirical model developed by Wang and Burbank (Reference 1).

[b]Available minimum liquid-to-gas ratio at BREC = 67.4 gal./10^3acfm.

TABLE 3. TYPICAL COAL ULTIMATE ANALYSIS BREC'S GREEN STATION UNIT NO. 2

	Percent by Weight	
	As Received	Dry Basis
Moisture	10.50	-
Carbon	58.86	65.77
Hydrogen	4.05	4.53
Nitrogen	1.23	1.37
Chlorine	0.05	0.06
Sulfur	3.53	3.94
Ash	14.44	16.13
Oxygen (diff.)	7.34	8.20
TOTAL	100.0	100.0

shown in Table 4 were also consistent with values measured by ETS, and were thus used as the basis for performing material balance calculations to establish required equipment sizes.

3.3 Retrofit Equipment Required

The existing lime system would be converted by utilizing as much existing equipment as possible. Only a ball mill and organic acid feed system will be added. The resulting simplified process flow diagram for the limestone preparation portion of the system is shown in Figure 1. No changes to the absorption portion of the system were required.

An existing equipment list of the BREC FGD system was prepared jointly by PPSI and BREC. A site visit was made by PPSI engineers to determine possible use of existing equipment in the limestone retrofit and to locate sites for new equipment construction.

Existing lime silos will be used for storage of coarse limestone; however, the lime slakers will not be operated. Three 50 TPH ball mills (one spare) will be housed in an electrically heated building near the existing silos. DBA or adipic acid storage and handling facilities will also be in the ball mill area. Conveyors have been priced to transport the crushed limestone from the storage silos to any of the three ball mill feeds. Supernate, tapped from a nearby 6 in. line, will be used to slurry the limestone, and the organic acid will also be added to the ball mill. A crane was priced for new ball addition to the ball mill.

Tanks, pumps, and agitators for the limestone/organic acid feed system were sized and priced based on maximum coal feed rate and maximum sulfur coal. It should be noted that costs for an electrical substation to provide power for the ball mills were not included since BREC felt that sufficient electrical capacity was available from existing equipment. If this substation were included, it would raise the capital costs by $52,000.

TABLE 4. COAL COMBUSTION CALCULATIONS

INPUT DATA

FUEL TYPE	COAL		
HEAT DUTY (MMBtu/hr)	4219.2	HIGHER HEATING VALUE (Btu/lb)	10548
TOTAL FLUE GAS W/O FLYASH (lb/hr)	5121000	EXCESS AIR SUPPLIED (%)	50.0
BOILER & AIR HEATER LEAKAGE (%)	0.0		
AMBIENT AIR TEMPERATURE (Deg F)	60	RELATIVE HUMIDITY (%)	0
PLANT ELEVATION NOT GIVEN - ASSUMED SEA LEVEL		AMBIENT AIR PRESSURE (PSIA)	14.696
FLUE GAS TEMPERATURE (Deg F)	320	FLUE GAS PRESSURE (In. WC)	0.00
STANDARD GAS CONDITIONS: 70 Deg F & 14.696 PSIA			

RESULTS USING NORMALIZED ULTIMATE ANALYSIS OF FUEL

FUEL FIRING RATE (lb/hr)	400000	TOTAL FLUE GAS VOLUME (ACFM)	1642370
THEORETICAL COMBUSTION AIR (ACFM)	698404		
ACTUAL COMBUSTION AIR (ACFM)	1047606	BOILER & AIR HEATER LEAKAGE (ACFM)	0
MOISTURE (lb water/lb dry gas)	0.0381	DRY FLUE GAS (lb/hr)	4932113
MOLECULAR WEIGHT (DRY/WET BASIS)	30.31 / 29.57	DENSITY OF FLUE GAS (lb/cu ft)	0.0520
SULFUR DIOXIDE CONC. (PPMD/PPMW)	2712 / 2549	HYDROGEN CHLORIDE CONC. (PPMD/PPMW)	35 / 33
SULFUR DIOXIDE CONC. (lb/MM Btu)	6.693	REQUIRED SO2 REMOVAL EFF. per NSPS (%)	90.00
TOTAL ASH IN FUEL (lb/hr)	57760		

FUEL ANALYSIS

FUEL CONSTITUENTS	WEIGHT PERCENT INPUT	WEIGHT PERCENT NORMALIZED
CARBON	58.860	58.860
HYDROGEN	4.050	4.050
NITROGEN	1.230	1.230
CHLORINE	0.050	0.050
SULFUR	3.530	3.530
WATER	10.500	10.500
ASH	14.440	14.440
OXYGEN	7.340	7.340
TOTALS	100.000	100.000

TOTAL FLUE GAS ANALYSIS

FLUE GAS CONSTITUENTS	FLOW RATE lb/hr	FLOW RATE MOLS/hr	% (W/O FLYASH) WEIGHT	% (W/O FLYASH) VOLUME
SULFUR DIOXIDE	28240	441	0.6	0.3
NITROGEN	3669310	131047	71.6	75.7
OXYGEN	371078	11596	7.2	6.7
CARBON DIOXIDE	863280	19620	16.9	11.3
HYDROGEN CHLORIDE	206	6	0.0	0.0
WATER	187749	10431	3.7	6.0
FLYASH	0			
TOTALS	5119863	173141	100.0	100.0

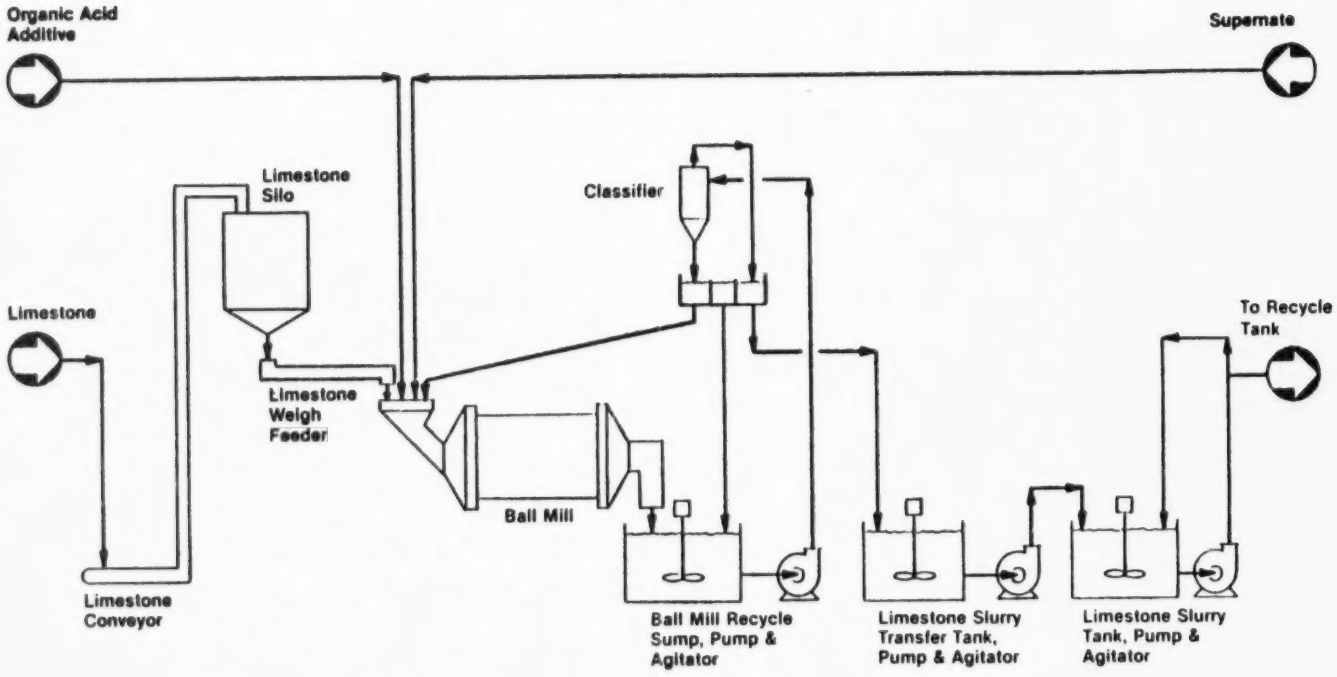

Figure 1. Limestone Preparation System

4.0 COSTS

In order to provide cost comparisons of various full-scale process alternatives, a base case of BREC was calculated using the parameters in Table 5. The typical operating load of 75 percent was obtained using BREC projected load demands through 1995 (see Figure 2).

Potential process alternatives considered in this evaluation are:

- Limestone - with adipic acid addition.
- Limestone - with DBA addition.

Direct and indirect capital requirements for new equipment are summarized in Table 6. Capital requirements are based on equipment sizing for 100 percent load and maximum sulfur. The ball mills are very conservative since the limestone work index supplied by BREC for a typical quarry was 12. Other quarries were not surveyed, but finding a work index of 10 or lower should not be difficult.

Operating costs shown in Table 7 were calculated based on typical conditions, since an unfair advantage for limestone/organic acid would result at maximum boiler load and coal sulfur.

By using a fixed capital recovery factor of 20 percent, the annualized costs of a limestone/organic acid retrofit and the existing lime system were compared. The results shown in Table 8 indicate a yearly savings of $2,642,000 for limestone with adipic acid and $3,153,000 for limestone with DBA.

TABLE 5. PARAMETERS ASSUMED FOR BASE CASE CALCULATIONS
BREC'S GREEN STATION UNIT NO. 2

Plant Load	=	205 MW
Module Gas Flow	=	1,525,000 lb/hr/tower
	or	383,000 acfm/module
Inlet SO_2 Concentration	=	2,550 ppm wet
Lime Stoichiometry	=	1.02
Lime Consumption	=	12 tons/hr
SO_2 Removal Required	=	90%
L/G (max.)	=	57
L/G (current)	=	41
CURRENT		
Recycle, Module A	=	14,900 gpm
Recycle, Module B	=	16,500 gpm
Operation	=	7,560 hr/yr
Cake Solids	=	42.0%

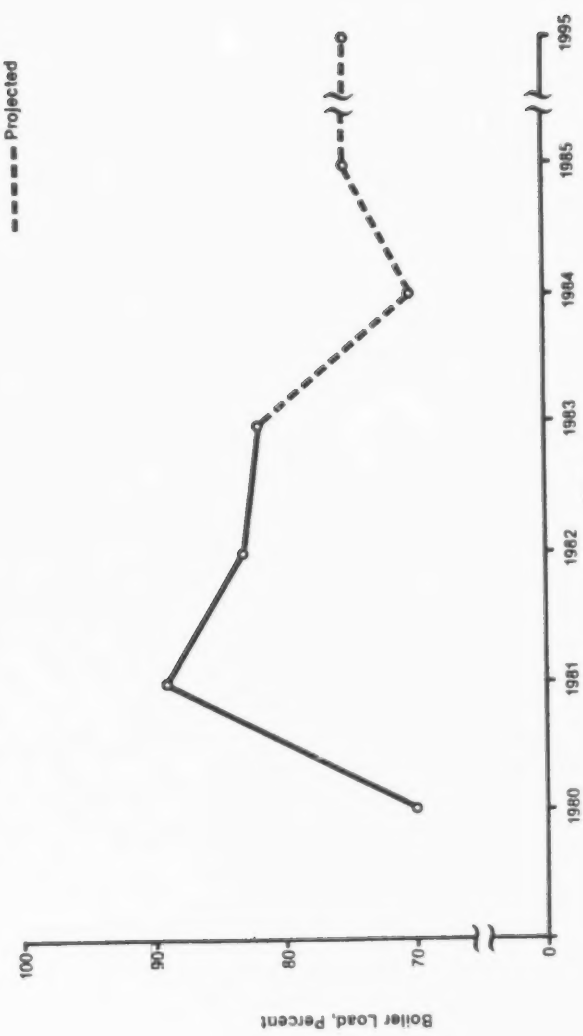

Figure 2. Green Station Operating Boiler Load; Actual and Projected

TABLE 6. COST ESTIMATE SUMMARY, BREC LIMESTONE/ORGANIC ACID RETROFIT
(1982 Cost Basis)

	Direct Investment ($1,000's)					
	Limestone/Adipic Acid			Limestone/DBA		
Description	Material	Labor	Total	Material	Labor	Total
Foundations, Site Prep.	50	50	100	50	50	100
Buildings	170	77	247	170	77	247
Structural Steel	94	56	150	94	56	150
Tank Heater				1	1	2
Process Tanks	4	2	6	24	8	32
Limestone Prep.	2,145	136	2,281	2,125	126	2,251
Pumps & Drives	18	4	22	18	4	22
Machinery	30	12	42	30	12	42
Insulation				6	6	12
Piping & Valves	105	65	170	105	65	170
Instrumentation	115	30	145	115	30	145
Electrical	173	108	281	173	108	281
Painting	1	4	5	1	4	5
Freight	25		25	25		25
Total Direct Costs	2,930	544	3,474	2,937	547	3,484
	Indirect Investment ($1,000's)					
Start-up Expenses		10	10		10	10
Field Expenses		20	20		20	20
Vendor Salary	100	20	120	100	20	120
Vendor Expenses	20		20	20		20
Vendor Overhead	100	20	120	100	20	120
Contingency	117	22	139	117	22	139
Vendor Profit @ 15%			585			587
Total Indirect Costs	337	92	1,014	337	92	1,016
Total Cost			4,488			4,500

TABLE 7. OPERATING COST SUMMARY - TYPICAL OPERATING CONDITIONS[a]
BREC LIMESTONE/ORGANIC ACID RETROFIT
(1982 Cost Basis)

	First Year Cost ($1,000's)		
	Lime	Limestone/Adipic	Limestone/DBA
Alkali			
Lime @ $60/ton	5,856		
Limestone @ $6.50/ton		1,222	1,222
Additive		931	418
Utilities			
Electricity @ $.01369/kWh	215	378	378
Total	6,071	2,531	2,018
Savings Compared to Lime System	-	3,540	4,053

[a]Typical Operating Conditions:
Firing Rate = 100 tons/hr per boiler
3.0% sulfur coal
Lime Stoichiometry = 1.02 mole CaO/mole SO_2 removed
Limestone Stochiometry = 1.05 mole $CaCO_3$/mole SO_2 removed

TABLE 8. BREC LIMESTONE/ORGANIC ACID RETROFIT PROJECTED FIRST YEAR COST DIFFERENTIAL FOR CONVERSION FROM LIME (1982 Cost Basis)

	Limestone/Adipic	Limestone/DBA
New Investment Annual Charge @ 20% ($1,000's)	898	900
New Annual Operating Costs ($1,000's)	-3,540	-4,053
Total Annual Savings		
$1,000's	2,642	3,153
mills/kWh	0.927	1.106

TABLE 9. CONVERTING UNITS OF MEASURE

To Convert From	To	Multiply By
Btu	J	1,055
scfm (60°F)	Nm^3/hr (0°C)	1.61
cfm	m^3/hr	1.70
°F	°C	(°F-32)/1.8
ft	m	0.305
ft^2	m^2	0.929
ft^3	m^3	0.0283
gal.	l	3.79
gal./10^3 acfm	l/m^3	0.134
gr	g	0.065
in.	cm	2.54
in. w.g.	mm Hg	1.87
lb	g	454
psi	kPa	6.895
ton	metric ton	0.907

REFERENCES

1. Wang, S. C., and D. A. Burbank, "Adipic Acid-Enhanced Lime and Limestone Testing at the EPA Alkali Scrubbing Test Facility - Final Report, Volume I. Basic Report," EPA-600/7-82-010a, NTIS PB 82-230624, March 1982, pp. 198-201.

NOTE: This paper is based on the Project Summary and Project Report entitled "Economics of Retrofitting a Lime Based FGD System for Operation with Limestone and Organic Acid;" this report, sponsored by EPA under Contract No. 68-02-3171, Work Assignment No. 78, with Radian Corporation, is in print.

TVA/OP/EDT-84/32

ECONOMIC EVALUATION OF LIMESTONE AND LIME FLUE GAS DESULFURIZATION PROCESSES FOR NEW SYSTEMS

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DISCLAIMER

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ABSTRACT

The preliminary-grade economics (accuracy: -15 percent, +30 percent) of various alternative limestone and lime flue gas desulfurization (FGD) processes are examined using the current design and economic premises established for the continuing series of economic evaluations performed by TVA for EPA. The economics are projected using the Shawnee lime/limestone computer model, which is based on long-term operating data from EPA's alkali scrubbing test facility at TVA's Shawnee Steam Plant near Paducah, KY. The capital investment for the base-case limestone scrubbing process (500 MW, 3.5 percent sulfur coal, 1979 NSPS, spray tower, forced oxidation, landfill) is $206/kW. The first-year and levelized annual revenue requirements are 10.59 and 15.09 mills/kWh, respectively. Costs for the equivalent limestone scrubbing process using a Turbulent Contact Absorber (TCA) are lower, while those for the venturi/spray tower absorber are higher. Forced-oxidation/landfill disposal has a lower capital investment than unoxidized/pond disposal for all options studied; however, the first-year and levelized annual revenue requirements are slightly higher for the forced-oxidation/landfill process for most coal applications. For the spray tower limestone process to achieve a specified SO_2 removal efficiency, it is more economical to increase the limestone stoichiometry and minimize the absorber liquid/gas ratio (L/G). The use of adipic acid or possibly dibasic acid (DBA) as an additive to enhance SO_2 removal in the limestone scrubbing process is an economically attractive option.

INTRODUCTION

This study evaluates the economics of limestone and lime wet scrubbing FGD processes that incorporate recent technological developments such as forced oxidation, refined absorber design and operating conditions, landfill disposal, and the use of additives to enhance scrubbing efficiency. Most of the processes are based on EPA-sponsored FGD evaluations at the Shawnee test facility. The economic data were obtained using a computer model developed

from the Shawnee test facility results.

The primary focus of this study is on absorber types, forced oxidation, and the use of additives. The three types of absorbers (Turbulent Contact Absorber (TCA), spray tower, and venturi/spray tower), the forced oxidation method (single loop, two loop, and bleedstream), and the type of additive (magnesium oxide (MgO) and adipic acid) are the main subjects. Landfill disposal is the primary disposal method used, but a range of pond disposal processes are included for comparison. All processes without forced (or natural) oxidation have pond disposal; all forced oxidation processes, except two, have landfill disposal.

Forced oxidation to gypsum has recently evolved into the mainstream of limestone and lime processes. Oxidation to nearly 100 percent gypsum is achieved, and the waste is readily dewatered to 85 percent or more solids. It is apparently tractable to landfill disposal without stabilization. Use of the gypsum for wallboard and cement manufacture is also a possibility.

The use of MgO to enhance SO_2 removal efficiency has been investigated for several years, mostly in lime processes, to increase dissolved alkalinity. Adipic acid, an organic additive which has recently received considerable attention, acts as a buffer that tends to maintain a liquid pH favorable to SO_2 absorption. It has been extensively evaluated by EPA as a means of improving limestone FGD performance.

PROCESSES EVALUATED

The processes evaluated consist primarily of combinations of (1) limestone or lime absorbent; (2) TCA, spray tower, or venturi/spray tower absorber; (3) pond disposal of sludge without forced oxidation or forced oxidation with landfill; and (4) no additive, and MgO or adipic acid additive. The alternatives of increased partial scrubbing (using adipic acid to enhance SO_2 removal efficiency) are also investigated.

RESULTS

The economic results consist of the capital investment in 1982 dollars, first-year annual revenue requirements (which include levelized capital charges) in 1984 dollars, and levelized (over a 30-year period) annual revenue requirements (which consist of levelized operating and maintenance costs as well as levelized capital charges). Summaries of these costs are shown in Table 1.

COMPARISON OF LIMESTONE PROCESSES WITH AND WITHOUT FORCED OXIDATION

Costs for comparable limestone processes without forced oxidation and with pond disposal and those with forced oxidation and landfill are shown in Table 2. Processes without forced oxidation and with pond disposal have higher capital investments than comparable processes with forced oxidation and landfill disposal, in all cases. The differences increase with increasing coal sulfur content. The higher capital investment is largely a result of disposal area construction costs, which are 2 to 3-1/2 times higher for ponds than for landfills. In contrast, the additional capital investment for forced oxidation and sludge dewatering is relatively small. Except for the 5.0 percent sulfur coal cases, first-year annual revenue requirements are higher for forced-oxidation processes because of the higher operating labor, maintenance, and overhead costs associated with the waste dewatering and landfill operations. Levelized annual revenue requirements are higher for the forced-oxidation processes. This is the result of the higher proportion of operating and maintenance costs in the total annual revenue requirements of the forced-oxidation processes.

ADDITIVES

Costs of comparable processes with and without additives are shown in Table 3.

TABLE 1. CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENTS

Plant size, MW	Coal sulfur, %	Absorber type[a]	Capital investment, 1982$		Annual revenue requirements, 1984$			
					First-year		Levelized	
			$10^6	$/kW	$10^6	mills/kWh	$10^6	mills/kWh
Limestone Processes Without Forced Oxidation[b]								
200	3.5	ST	62.7	314	15.8	14.39	21.7	19.72
500	0.7	ST	73.2	146	18.1	6.56	24.5	8.91
500	2.0	ST	90.4	181	23.3	8.46	32.1	11.68
500	3.5	ST	112.6	225	28.8	10.47	39.6	14.42
500	5.0	ST	123.3	247	31.5	11.44	43.3	15.73
500	3.5	TCA	107.8	216	27.6	10.03	38.0	13.81
500	3.5	V-ST	121.7	243	31.3	11.37	43.1	15.68
1,000	3.5	ST	192.1	192	48.9	8.89	67.2	12.23
Limestone Processes With Forced Oxidation[c]								
200	0.7	ST	44.4	222	12.3	11.15	17.3	15.76
200	2.0	ST	52.1	260	14.7	13.40	21.0	19.10
200	3.5	ST	59.1	295	16.7	15.21	23.9	21.69
200	5.0	ST	62.6	313	17.8	16.17	25.4	23.08
500	0.7	ST	70.4	141	18.8	6.84	26.3	9.56
500	2.0	ST	86.6	173	24.2	8.80	34.4	12.49
500	3.5	ST	103.1	206	29.1	10.58	41.5	15.08
500	5.0	ST	109.2	218	31.2	11.36	44.7	16.25
500	3.5	TCA	98.5	197	27.8	10.09	39.5	14.37
500	3.5	V-ST	109.2	218	30.8	11.20	43.9	15.96
500[d]	3.5	ST	94.4	189	25.8	9.40	36.4	13.25
1,000	0.7	ST	109.2	109	28.3	5.15	39.2	7.13
1,000	2.0	ST	151.7	152	41.1	7.47	57.7	10.50
1,000	3.5	ST	177.2	177	49.3	8.97	69.9	12.71
1,000	5.0	ST	187.0	187	53.0	9.63	75.5	13.73
Limestone Processes With Forced Oxidation and Additive Adipic Acid[e]								
200	3.5	ST	54.4	272	15.3	13.95	21.9	19.86
500	0.7	ST	61.4	123	16.4	5.97	23.0	8.36
500	2.0	ST	79.1	158	21.6	7.86	30.4	11.07
500	3.5	ST	92.8	186	25.8	9.38	36.6	13.31
500	5.0	ST	97.9	196	27.7	10.07	39.5	14.36
500	3.5	TCA	90.4	181	25.4	9.23	36.1	13.12
500	3.5	V-ST	96.3	193	27.0	9.81	38.3	13.93
500	3.5[f]	ST	100.7	201	26.8	9.73	37.3	13.58
500	3.5[g]	ST	92.8	186	25.7	9.35	36.4	13.25
1,000	3.5	ST	155.5	155	42.8	7.78	60.4	10.99
MgO								
500	0.7	ST	69.1	138	18.4	6.70	25.8	9.37
500	2.0	ST	86.6	173	23.8	8.66	33.6	12.23
500	3.5	ST	102.5	205	28.8	10.49	41.1	14.93
500	5.0	ST	107.9	216	30.7	11.18	43.9	15.98
500	3.5	TCA	98.2	196	27.7	10.09	39.5	14.38
500	3.5	V-ST	113.1	226	31.6	11.50	44.9	16.33
500	3.5	ST	111.2	222	28.5	10.36	39.2	14.27
Lime Processes Without Forced Oxidation[b]								
500	0.7	ST	60.7	121	15.3	5.56	20.9	7.62
500	2.0	ST	79.3	159	21.8	7.92	30.7	11.18
500	3.5	ST	96.2	192	28.0	10.19	40.3	14.66
500	5.0	ST	105.3	211	31.6	11.51	46.0	16.71
500	3.5	TCA	94.6	189	27.7	10.09	40.0	14.55
500	3.5	V-ST	103.2	206	29.9	10.88	43.0	15.63
Lime Processes With Forced Oxidation[c]								
500	3.5	ST	88.9	178	28.5	10.35	42.1	15.31
500	3.5	TCA	86.2	172	27.9	10.16	41.5	15.07
500	3.5	V-ST	92.7	185	29.6	10.77	43.8	15.92
Lime Processes With Forced Oxidation and MgO[c]								
500	3.5	ST	89.6	179	28.6	10.40	42.3	15.37
500	3.5	TCA	90.4	181	28.7	10.45	42.4	15.43
500	3.5	V-ST	96.8	194	30.3	11.01	44.5	16.18

[a] ST=spray tower; TCA=Turbulent Contact Absorber; V-ST venturi/spray tower.
[b] Pond disposal in all cases.
[c] Landfill disposal in all cases.
[d] No flue gas reheat.
[e] Landfill disposal, except as noted.
[f] Pond disposal.
[g] Dibasic acid.

TABLE 2. COST COMPARISON OF PROCESSES WITH AND WITHOUT FORCED OXIDATION

Process[a]	Capital investment, 1982 $/kW	Annual revenue requirements, 1984 mills/kWh	
		First-year	Levelized
0.7% Sulfur Coal[b]			
No forced oxidation, pond	146.5	6.56	8.91
No forced oxidation, landfill	140.7	6.84	9.56
2.0% Sulfur Coal			
No forced oxidation, pond	180.8	8.46	11.68
Forced oxidation, landfill	173.1	8.80	12.49
3.5% Sulfur Coal			
No forced oxidation, pond	225.1	10.47	14.42
Forced oxidation, landfill	206.1	10.58	15.08
5.0% Sulfur Coal			
No forced oxidation, pond	246.7	11.44	15.73
Forced oxidation, landfill	218.3	11.36	16.25

[a] 500-MW, limestone spray tower processes, natural oxidation to 95% $CaSO_4 \cdot 2H_2O$ for 0.7% sulfur coal cases, in-loop forced oxidation to 95% $CaSO_4 \cdot 2H_2O$ for all other cases.

[b] With low sulfur coal, forced oxidation is not required to achieve high (>90%) oxidation of sulfite to sulfate.

TABLE 3. COST COMPARISON OF PROCESSES WITH AND WITHOUT ADDITIVES

	Capital investment, 1982 $/kW	Annual revenue requirements, 1984 mills/kWh	
		First-year	Levelized
Landfill Disposal[a]			
0.7% sulfur coal			
No additive	140.7	6.84	9.56
Adipic acid	122.8	5.97	8.36
MgO	138.2	6.70	9.37
2.0% sulfur coal			
No additive	173.1	8.80	12.49
Adipic acid	158.2	7.86	11.07
MgO	171.8	8.72	12.39
3.5% sulfur coal			
No additive	206.1	10.58	15.08
Adipic acid	185.5	9.38	13.31
MgO	205.0	10.49	14.93
5.0% sulfur coal			
No additive	218.3	11.36	16.25
Adipic acid	195.9	10.07	14.36
MgO	215.8	11.18	15.98
Pond Disposal[b]			
3.5% sulfur coal			
No additive	225.1	10.47	14.42
Adipic acid	201.4	9.73	13.58
MgO	222.4	10.36	14.27

[a] 500-MW, limestone spray tower processes, natural oxidation for all 0.7% sulfur coal cases, in-loop forced oxidation for no additives and adipic acid cases, bleedstream forced oxidation for all MgO processes.

[b] 500-MW, limestone spray tower processes without forced oxidation.

MgO

The use of MgO as an additive in the limestone process has no significant effect on capital investment. The equipment needed to handle the relatively small volumes of additive does not materially affect equipment costs. By improving SO_2 removal efficiency and absorbent utilization, the use of MgO slightly decreases gas handling, SO_2 absorption, and disposal costs. The effect is thus more pronounced at higher coal sulfur contents. For 3.5 percent sulfur coal, the reduction in capital investment is less than 1 percent, however.

The effect of MgO on annual revenue requirements of forced-oxidation processes is similar to the effect on capital investment. The cost of the additive is minimal (less than 1 percent of total annual revenue requirements). The improved SO_2 removal efficiency and absorbent utilization, expressed primarily as reduced absorbent and electricity costs, more than offset the cost for MgO. For 3.5 percent sulfur coal, the cost reduction is less than 1 percent.

Adipic Acid

In contrast to the results with MgO addition, the use of adipic acid appears to be an economically attractive additive for all applications considered in this study. Capital investment reductions range from nearly 13 percent for 0.7 percent sulfur western subbituminous coal to about 10 percent for high-sulfur coal. The additional equipment needed to handle the small quantities of adipic acid does not have an appreciable effect on equipment costs. However, by improving SO_2 removal efficiency and absorbent utilization, the use of adipic acid decreases the capital investments for the feed preparation, gas handling, SO_2 absorption, solids separation, and disposal area costs. The primary reasons for the more dramatic reductions in costs are the assumptions that with adipic acid the limestone stoichiometry can be reduced by 24 percent (1.07 versus 1.40) and that

95 percent SO_2 removal can be achieved, allowing partial flue gas bypass to reduce both the reheat requirements and the size of the absorbers. (Note that vendors are quoting stoichiometries as low as 1.10 on bids for processes without additives; this would obviously result in a lower cost advantage for the adipic-acid-enhanced process.)

The effect of adipic acid addition on both the first-year and the levelized annual revenue requirements for the forced-oxidation processes is similar to its effect on capital investment: it decreases costs 11-13 percent, depending on the coal sulfur level. Most of this cost reduction is due to a savings in capital charges. Other reductions (e.g., for electricity, absorbent, maintenance, and overhead) also contribute to a significant degree. For forced oxidation, the cost of adipic acid is minimal (less than 1 percent of annual revenue requirements for all except 5.0 percent sulfur coal); and the improved SO_2 removal efficiency and absorbent utilization, expressed primarily as reduced absorbent and electricity costs, more than offset the cost for adipic acid.

In processes without forced oxidation, the annual revenue requirements of processes using additives are lower than those of the same processes without additives, but the difference is less because of the additional additive lost in the low-solids wastes. The cost of the additive in these cases is about the same as the absorbent cost. Except for the additional additive costs, the other cost reductions remain unchanged.

ABSORBER TYPE

Cost comparisons for limestone processes with different types of absorbers are shown in Table 4 for limestone processes at the 500-MW, 3.5 percent sulfur coal conditions. The capital investment relationships of the processes with TCA's, spray towers, and venturi/spray towers remain the same whether forced oxidation or additives are used. For all, the TCA process has the lowest capital investment, and the venturi/spray tower has

TABLE 4. COST COMPARISON BY ABSORBER TYPE

Process[a]	Capital investment, 1982 $/kW	Annual revenue requirements, 1984 mills/kWh	
		First-year	Levelized
Without Forced Oxidation (Pond Disposal)			
Spray tower	225.1	10.47	14.42
TCA	215.6	10.03	13.81
Venturi/spray tower	243.4	11.37	15.68
With Forced Oxidation (Landfill Disposal)			
Spray tower	206.1	10.58	15.08
TCA	197.0	10.09	14.37
Venturi/spray tower	218.4	11.20	15.96
With Forced Oxidation and Adipic Acid (Landfill Disposal)			
Spray tower	185.5	9.38	13.31
TCA	180.9	9.23	13.12
Venturi/spray tower	192.7	9.81	13.93

[a] 500 MW, 3.5% S.

the highest.

The major capital investment advantage of the TCA process lies in lower costs for SO_2 absorption because of the smaller absorber and lower L/G. Processes using the larger, lower-pressure-drop spray tower have higher capital investments for SO_2 absorption than the TCA processes, which are only partially offset by the lower capital investment for fans. The processes using the venturi/spray tower absorbers have a higher limestone utilization rate and thus slightly lower capital investments for feed preparation, solids dewatering, and disposal. This is more than offset by the higher capital investments for gas handling and SO_2 absorption. The same cost relationships also prevail in annual revenue requirements, largely because of costs based on capital investments such as maintenance and capital charges. In costs other than maintenance and capital charges, there is little difference in annual revenue requirements between the processes with TCA's, spray towers, or venturi/spray towers.

SPRAY TOWER ABSORBER DESIGN CONDITIONS

Since the superficial gas velocity through the absorber, the pH of the slurry and the SO_2 removal efficiency are normally determined by the technology or the performance requirements, the absorber L/G ratio and stoichiometry are the only remaining design conditions that can be optimized. Thus it becomes an economic tradeoff between the costs associated with slurry recirculation pumps (L/G) and the costs for the limestone preparation equipment and the waste disposal area (stoichiometry). Increasing the limestone stoichiometry and decreasing the absorber L/G is the more economical option. Increasing the stoichiometry from 1.1 to 1.4 mol $CaCO_3$/mol (SO_2 + 2HCl) absorbed results in a decrease in capital investment of 14-18 percent. Levelized annual revenue requirements decrease 16-18 percent.

PARTIAL VERSUS FULL SCRUBBING

For lower sulfur coals where relatively modest SO_2 removal efficiencies are required, two alternative designs are possible: full scrubbing at the required removal efficiency, or scrubbing part of the flue gas at a high removal efficiency (e.g., 90 percent); and bypassing the rest of the flue gas around the absorber to reduce or eliminate the need for stack gas reheat. The costs of full and partial scrubbing for a range of coal sulfur levels are based on a 500-MW limestone process with a spray tower absorber, forced oxidation, and landfill disposal. Partial scrubbing is based on 90 percent SO_2 removal in the absorber. In terms of capital investment, it is less expensive to partially scrub the flue gas from coals with sulfur contents below about 1.5 percent. For 0.7 percent sulfur western subbituminous coal, partial scrubbing saves about $10/kW in capital investment (a 6.5 percent reduction) and about 1.2 mills/kWh in annual revenue requirements (an 11.2 percent reduction). For 2.0 percent sulfur coal, full scrubbing saves $4.0/kW in capital investment over partial scrubbing, but the annual revenue requirements are essentially the same in both cases.

FULL-SCRUBBING PROCESSES AT 85 PERCENT, 90 PERCENT, and 95 PERCENT SO_2 REMOVAL

As the required SO_2 removal efficiency increases from 85 percent to 95 percent, the economic advantage of the adipic-acid-enhanced process increases. At 85 percent removal, the adipic-acid-enhanced process is about 9 percent lower in capital investment and nearly 12 percent lower in annual revenue requirements. At 95 percent removal, the corresponding reductions are more than 15 percent for both capital investment and annual revenue requirements.

LIMESTONE AND LIME PROCESSES

Limestone and lime process costs are compared for different coal sulfur contents and by process type in Table 5. The processes using lime maintain a substantial advantage in capital investment for all conditions. The advantage decreases slightly with increasing coal sulfur content. At the same coal sulfur content, the capital investment advantage of the lime processes with variations such as forced oxidation and additives remains relatively constant (about 15 percent lower than the corresponding limestone processes).

Coal sulfur content has an important effect on relationships of the annual revenue requirements. The lime process has lower annual revenue requirements for the low-sulfur coal cases, but the advantage decreases with increasing coal-sulfur content because of increasing absorbent costs. The different relationship in first-year and levelized annual revenue requirements occurs because levelizing increases operating and maintenance costs, which are a significantly lower portion of annual revenue requirements for the limestone process.

REHEAT

Since in-line steam reheat involves significant expenditures in both capital investment and annual revenue requirements and since benefits of reheat sometimes are not apparent, some utilities have specified FGD systems without reheat. In comparing two 500-MW, 3.5 percent sulfur coal, limestone forced-oxidation processes, the capital investment without reheat is more than 8 percent lower ($18/kW) than with reheat. The annual revenue requirements are about 12 percent lower without reheat. These lower costs are due partly to the elimination of steam consumption and partly to the reduction in maintenance, overhead, and capital charges because of the lower capital investment for the case without reheat.

TABLE 5. COST COMPARISON OF LIMESTONE AND LIME PROCESS

Process	Capital investment, 1932 $/kW	Annual revenue requirements, 1984 mills/kWh	
		First-year	Levelized
Spray Tower[a]			
0.7% sulfur coal			
Limestone	146.5	6.56	8.91
Lime	121.5	5.56	7.62
2.0% sulfur coal			
Limestone	183.4	8.46	11.61
Lime	158.6	7.92	11.81
3.5% sulfur coal			
Limestone	225.1	10.47	14.42
Lime	192.3	10.19	14.66
5.0% sulfur coal			
Limestone	246.7	11.44	15.73
Lime	210.6	11.51	16.71
Venturi / Spray Tower[b]			
Without forced oxidation (pond)			
Limestone	*243.4*	*11.37*	*15.68*
Lime	*206.3*	*10.88*	*15.63*
Forced oxidation (landfill)			
Limestone	*218.4*	*11.20*	*15.96*
Lime	*185.4*	*10.78*	*15.92*
With forced oxidation and MgO[c] (landfill)			
Limestone	*226.1*	*11.50*	*16.33*
Lime	*193.5*	*11.01*	*16.18*

[a] *500-MW, without forced oxidation, pond disposal.*
[b] *500-MW, 3.5% sulfur coal.*
[c] *Bleedstream forced oxidation.*

CONCLUSIONS

Processes with forced oxidation and landfill disposal have lower capital investments than processes without forced oxidation and pond disposal because the aggregate costs of forced oxidation and dewatering equipment and landfill construction are small, compared with the cost of pond construction. First-year annual revenue requirements are slightly higher for forced-oxidation processes under most conditions evaluated. At large waste volumes, however, the first-year annual revenue requirements for processes with forced oxidation are slightly lower. All processes without forced oxidation have slightly lower levelized annual revenue requirements.

The use of additives reduces both the capital investment and annual revenue requirements of limestone processes. For processes with forced oxidation and landfill disposal, the effects of MgO are marginal; but with adipic acid, more substantial reductions of 10-13 percent are obtained in both capital investment and annual revenue requirements. The cost reductions are reflected in capital investment by reduced equipment sizes and smaller waste volumes; and in annual revenue requirements, by reduced capital charges (the major factor) and other costs related to capital investment and by reduced absorbent and electricity costs. The larger cost reductions obtained with adipic acid are largely a result of a substantial improvement in limestone utilization. The costs related to the use of additives are a very small part of the total costs, and both capital investment and annual revenue requirements are very insensitive to changes in the cost and consumption rate of the additives or to the use of lower cost substitutes. The higher removal efficiencies attainable with adipic acid also permit the use of the more economical partial scrubbing with bypass over a wider range of coal sulfur contents. For processes that use pond disposal, the use of additives has the same effect on capital investment as it does for processes using landfill disposal, but the cost reduction in annual revenue requirements is reduced by about half because of the additional additive loss in the low-solids waste.

Processes using TCA's are about 5 percent less expensive than processes using spray towers and 12 percent less expensive than processes using venturi/spray towers because of the smaller size and lower L/G ratio of the high-pressure-drop TCA, assuming that maintenance costs are the same percentage of equipment costs for all the processes. Changes of 15 percent in maintenance costs for either process make the annual revenue requirements of processes using TCA's and spray towers equivalent.

Economically, the optimum absorber design favors higher limestone stoichiometries and lower L/G ratios. For the same SO_2 removal efficiency, a 1.4 stoichiometry and a low L/G is 14-18 percent lower in capital investment and annual revenue requirements than a design with a 1.1 stoichiometry and a high L/G ratio.

Partial scrubbing at 90 percent removal efficiency with bypass of some of the flue gas is more economical at coal sulfur contents below about 2 percent because the bypassed flue gas reduces reheat costs.

All lime processes evaluated have lower capital investments than limestone processes. They also have lower annual revenue requirements at lower coal sulfur contents. However, lime processes are more sensitive to absorbent costs, and to levelizing since operating and maintenance costs constitute a larger percentage of the total annual revenue requirements. As a result, lime process annual revenue requirements increase more rapidly with coal sulfur content than those of limestone processes, particularly if levelized annual revenue requirements are compared. At a coal sulfur content of about 4.5 percent, the first-year annual revenue requirements of lime and limestone processes are equal; at a coal sulfur content of about 3.0 percent, the levelized annual revenue requirements are also equal. As the coal sulfur content further increases, limestone processes become increasingly more economical.

NOTE: This paper is based on the report, "Economic Evaluation of Limestone and Lime Flue Gas Desulfurization Processes," EPA-600/7-83-029, May 1983, NTIS No. PB 84-133644.

APPENDIX A

SUMMARY OF TOUR

SUMMARY OF SAN MIGUEL FGD SYSTEM OPERATIONS DURING THE INDUSTRY BRIEFING TOUR

INTRODUCTION

On July 19, 1984, EPA sponsored an Industry Briefing at the Marriott Hotel in San Antonio, Texas. The focus of this briefing was to disseminate information on the results of a test program conducted at San Miguel Electric Cooperative's Unit 1 FGD system. The test program involved adding organic acids to San Miguel's FGD system in an attempt to reduce process operating costs. The program was significant since it represented the first use of organic acids to reduce costs and it was successful; use of organic acids at San Miguel has resulted in an estimated reduction in 1984 annual operating costs of over $600,000. Because of the program's success and the potential for significant savings at other FGD systems, EPA sponsored the Industry Briefing to provide test results to utilities in a timely manner.

The San Miguel Industry Briefing consisted of two parts. In the morning, attendees were taken on a tour of the San Miguel Unit 1 FGD system. Then in the afternoon, papers were presented summarizing the results of the San Miguel test program and other recent developments concerning organic-acid-enhanced limestone scrubbing.

The discussion which follows summarizes how the San Miguel FGD system was operating during the tour. The discussion also describes changes in process operation which were made possible as the result of organic acid addition to the FGD system.

TOUR DESCRIPTION

TOUR SUMMARY

Five groups of 45 people each were given a tour of the San Miguel FGD system as part of the Industry Briefing. Because of time constraints, only the FGD system was included on the tour. The areas of the FGD system which were seen and discussed on each tour were:

(1) Reagent Preparation Area
(2) Scrubber Area
(3) Organic Acid Feed Area
(4) Dewatering and Disposal Area

The operation of each of these areas, as observed during the San Miguel tour, is summarized below.

GENERAL PLANT OPERATION

The San Miguel plant is a mine-mouth facility. It consists of a single 410 MW lignite-fired boiler equipped with an electrostatic precipitator for particulate control and an FGD system for controlling SO_2 emissions. Since the plant is a new base-loaded unit (commercial operation began in January 1982), it often operates above its design capacity of 410 MW's. This was the case during the Industry Briefing tour.

The FGD system is a Babcock and Wilcox designed wet limestone scrubber. It consists of limestone reagent preparation and storage facilities, four 33 percent scrubber modules, an organic acid feed system, and equipment for dewatering and disposal of waste sludge.

TOUR STOPS

REAGENT PREPARATION

Limestone reagent for the FGD system is received in 100 ton rail cars as 3/4" x 3/8" stone. This coarse stone is then processed in the reagent preparation area of the San Miguel FGD system which consists of limestone unloading and storage facilities, a ball mill for limestone grinding, hydrocyclones for ground limestone classification, and a limestone slurry storage tank which supplies reagent to the scrubber modules.

The reagent preparation system is designed to product a 30 weight percent solids slurry of finely ground limestone for use in the FGD system. Coarse limestone is ground wet in a 48 ton/hr ball mill and classified in a bank of 4 inch diameter hydrocyclones. The water used for grinding is supplied from the thickener overflow tank which is located in the dewatering area. San Miguel's mill circuit was designed to product 95 percent past 325 mesh limestone and actual product size averages 90 percent past 325 mesh. While the tour took place, the mill and classification circuit was in operation at full capacity. With organic acid addition, the mill typically operates 10 to 12 hours per day while prior to organic acid addition at San Miguel, the mill operated for 16 to 18 hours per day. The reduction in operating time corresponds to a reduction in limestone stoichiometry which was made possible with organic acid addition.

ABSORBER MODULES

General

The scrubber area of the San Miguel FGD system consists of four identical modules each sized for 33 percent gas flow at full load. Thus during typical operation (as was the case during the tour), only three of the modules are in operation. Each of the scrubber modules has separate quench and absorber towers for processing the flue gas.

1,500,000 acfm of flue gas from the 410 MW lignite-fired boiler is processed by the three operating scrubber modules. Since the time DBA addition became part of normal process operation, about 10,000 acfm of flue gas is bypassed to help reheat the flue gas leaving the FGD system. The flue gas which is scrubbed first enters a venturi-type quench tower. At this point the gas is about 320°F and contains approximately 2,500 ppm SO_2. The gas passes down through the quench tower where it is saturated and some of the SO_2 is removed. It then turns 180° and passes up through the tray-type absorber tower where much of the remaining SO_2 is removed. With DBA, the typical SO_2 removal for each quench-absorber module is over 90 percent.

Scrubbed flue gas which leaves the absorber passes through a horizontal mist eliminator and then enters a reheat chamber. In the reheat chamber the scrubbed, saturated flue gas is mixed with bypassed flue gas and heated air which reheats it from 135° to 170°F. This gas is then discharged through the Unit 1 stack.

Slurry in the scrubber loop is recirculated to the quench and absorber towers from separate tanks and limestone reagent is added to both tanks on pH control. Slurry from the absorber recirculation tank overflows to the quench loop and a bleed stream from the quench loop is pumped to the thickener for dewatering and eventual disposal. Makeup water which consists of a mixture of thickener overflow, cooling tower blowdown, and ash pond water, is added to each scrubber loop to maintain the solids density in the recirculating slurry at about 15 weight percent. Makeup water also enters the scrubber loop as mist eliminator wash water.

As the time of the tour, DBA or dibasic acid was the organic acid which was added to the FGD system. The concentration of this material in the recirculating slurry was 500 to 600 ppm which is the range identified as economically optimum during the test program. The DBA was present in the slurry recirculated in both the quench and absorber loops.

Quench Loop

The quench loop operates independently from the absorber loop in each scrubber module. Quench loop L/G is about 11 gal/macf and with DBA addition to the system, it is possible to maintain the recirculating slurry pH at 5.4. This is substantially lower than the operating pH range of 6.0 to 6.5 observed prior to DBA addition. It is operation at the lower pH which is primarily responsible for the improvement in limestone utilization which has been observed at San Miguel.

Absorber Loop

The absorber has one level of spray nozzles and two perforated plate trays. Gas/liquid contact in the absorber loop provides the majority of SO_2 removal in the system. L/G in the absorber is about 44 gal/macf and it operates at a pH of 5.8. This is also lower than the pH's of 6.2 or greater which were typical prior to DBA addition. It is believed that operation at the lower pH in the absorber loop (which corresponds to about 85 percent limestone utilization) is responsible for the reduction in absorber and mist eliminator plugging which has been observed at San Miguel since the start of the organic acid addition program.

ORGANIC ACID FEED SYSTEM

The organic acid feed system at San Miguel consists of a 30,000 gallon, insulated, stainless steel tank for storage of organic acid solutions, a small gear pump for feeding acid to the FGD system, and rotameters and hand operated valves for metering acid to the system. The storage tank is equipped with a fitting to allow direct unloading of tanker trucks of acid. Plant air is used to pressurize the tanker and force acid into the storage tank.

San Miguel's organic acid feed system has been through several changes since the organic acid addition program started. Initially the system was configured to feed organic acid directly to the absorber recirculation tanks of each module. However, shortly after the start of the test program, corrosion problems were encountered in the feed system and the organic acid feed lines had to be replaced. To minimize down-time, San Miguel replaced the individual module feed lines with two insulated lines which feed organic acid to the thickener overflow tank in the dewatering area of the FGD system. Since thickener overflow is returned to the recirculation tanks, the organic acid is indirectly fed to the absorber and quench loops. This feed system configuration was in operation during the tour but San Miguel still considers it a temporary fix.

DEWATERING/DISPOSAL

The dewatering and disposal area of the FGD system consists of two 110 ft diameter thickeners, thickener overflow and underflow storage tanks, three rotary drum vacuum filters, and two pug mills. In this area, a bleed stream from each operating quench module is first processed in the thickeners to concentrate waste solids. The solids-free, thickener overflow is sent to a storage tank and eventually returned to the scrubber modules. The 35 weight percent solids thickener underflow is held in a tank and processed by the rotary vacuum filters which further concentrate the waste solids to about 60 weight percent. The filtrate from this operation is sent to the thickener overflow tank while the filter cake is sent to the pug mills. There it is blended with fly ash to a 90 weight percent solids mixture which is returned to the lignite mine for disposal.

At the time of the tour, the vacuum filters and mills were not in operation. This is typical at San Miguel. Normally the dewatering equipment is only operated for two shifts a day since the equipment was designed with excess capacity. This area of the FGD system has also been affected by the use of organic acid. Prior to the test program, San Miguel

operated all three vacuum filters for 18 hours per day. With organic acid addition and the accompanying reduced limestone stoichiometry, the dewatering properties of the solid waste have improved and the waste volume has been reduced. As a consequence, only two of the vacuum filters are not operated 16 to 18 hours per day.

APPENDIX B

SAN MIGUEL ELECTRIC COOPERATIVE, INC.

FACT SHEET

SAN MIGUEL ELECTRIC COOPERATIVE, INC.
FACT SHEET

A. GENERAL

San Miguel Unit No. 1 is a typical "semi-outdoor" power station with boiler, turbine generator, and high pressure heater located in an open structure. The area below the turbine deck is enclosed and houses the turbine and cycle auxiliaries, electonics room, instrument shop, and the makeup demineralizer equipment.

Also located on the 330 acre plant site is a cooling tower and several retention ponds. The plant currently is classed as a "zero discharge" installation and has a nine acre retention pond to collect runoff from contaminated power plant areas. There is a twenty-one acre well storage pond and two thirteen acre ash water storage ponds which also serve for emergency storage of boiler ash.

B. WATER

Water supplies for plant operations are obtained from the Carrizo sands aquifier located beneath the plant site area. This aquifier is tapped by a system of interconnected wells, pumps and storage ponds.

C. TRANSPORTATION

Transportation to the plant is provided by a system of public highways including State Highways 16 and 281, Farm to Market Road FM-140 and Ranch Road 3387. Rail service is provided by Missouri Pacific to a siding which is connected by a San Miguel 13.5 mile long railroad spur system terminating at the plant site.

D. TURBINE GENERATOR

The turbine-generator is a Westinghouse tandem compound two flow reheat condensing steam turbine, 3600 RPM with 31" last stage blades. The generator is conventionally hydrogen cooled rated at 450 MVA at 60 psig H_2 and has a brushless excitation system. The turbine is guaranteed at 410 MW at 2400 psig/1000 F, 2-1/2 in. Hg. absolute exhaust when operating with the designed seven heater cycle, no makeup, no process steam extraction, and no boiler spray water takeoff but with a condensing boiler feed pump turbine in the cycle at the crossover extraction point. Comparable maximum expected load at 5% overpressure is 448 MW.

The cycle is designed for extraction of steam for SO2 scrubber outlet reheating and air preheating, and a relatively high makeup

rate (3.8% of throttle) necessitated by the high rate of soot blowing steam use and associated boiler continuous blow-down. These result in reduction in unit output to 401 MW at maximum guaranteed flow and 437 MW at maximum expected. Of the 11 MW reduction in maximum, 8 MW is a result of scrubber reheating.

E. BOILER

The steam generator is a Babcock & Wilcox natural circulation, balanced draft, drum-type boiler. The maximum continuous rate is 3,055,000 pounds of steam per hour at 2600 psi, 1005 F with a reheat of 2,705,000 pounds per hour at 575 psig, 1005 F.

The steam generator and its appurtenances, the electrostatic precipitator and the flue gas desulfurization system was purchased from, and erected by, B&W.

Seven B&W MPS-89 roller-type pulverizers are used. Lignite exhibits grindability values which are heavily dependent on moisture content, therefore, the pulverizers were selected based on the two highest moisture contents of samples analyzed. The pulverizers were conservatively selected and full boiler output can be achieved with any two pulverizers out of service. However, the plant normally operates with only one out of service. Preheated primary air dries the lignite in the pulverizers and transports it to the burners. For mechanical reasons, 700°F was specified as a limit for primary air temperature.

The high ash levels, as well as high quartz and silica levels, result in an abrasive lignite. Because of this and the high throughputs from the mill, a ceramic classifier cone is used. The piping from the pulverizers was sized to permit maximum turndown and velocities to reduce the effect of erosion. Operating experience has shown that ceramic elbows must be installed within the next twelve (12) months at the present rate of coal pipe repairs.

There are 42 dual register burners arranged 3 high by 6 wide on the front wall and 4 high and 6 wide on the rear wall. The heat input per burner at rated load is 105 million Btu with all burners in service. Each pulverizer feeds one horizontal row of six burners. The windowbox is compartmented into separate burner rows. Air foils and dampers are provided for measuring and controlling secondary air to each row of six burners. Distribution of air to each burner within these rows is achieved by setting of the individual burner registers. NO_x emissions from the boiler are controlled below 0.7 pounds per million Btu input by inherent burner designs controlling the excess air to the burners and conservative furnace and burner design.

The furnace is approximately 67 feet wide and 45 feet deep. The top of the boiler support steel structure is 291 feet above grade.

The unit is of the open furnace design with no division walls or platens. The tube walls are gas-tight welded membrane construction made up of 2-1/2" O.D. tubes on 3" centers. The furnace is cooled by natural circulation. Internally ribbed tubes are installed in the higher heat input zones for protection against tube failures caused by loss of nucleate boiling.

Analysis, burn tests and reviewed experience in other parts of the world where low grade fuels have been fired, resulted in the following design parameters:

No division walls or platens in the furnace
Volume heat release - 7,500 Btu per cubic foot per hour
Area heat release - 85,000 Btu per square foot per hour
Plan heat release - 1,465,000 Btu per square foot per hour
Burner zone heat release - 465,000 Btu per square foot per hour
Maximum furnace exit gas temperature - 1850F, HVT
Maximum gas velocity - 45 ft. per second

In addition, minimum side spacing of tubes in the convection passes is limited to 24 inches at temperatures between 1850 to 1650 F, HVT; 12 inches at 1650 to 1500 F, HBT; 9 inches at 1500 1150 F, HVT; and 4-1/2 inches thereafter.

Flue gas for this system is taken from the economizer hopper, passing through ash separating chambers and then through mechanical dust collectors. Recirculation through the furnace hopper is used for reheat steam temperature control at partial loads. Recirculation through upper furnace ports controls furnace exit gas temperature at high loads. The system was designed to achieve low wear rates on the two full capacity fan rotors. However, this has not proven to be the case based on the current rate of wear. Studies are currently under way to determine the feasibility of operations without gas recirculation.

The superheater includes the convection pass sidewalls, roof tubes, horizontal primary surface and secondary pendant surface. The front bank of pendant secondary superheater tubes is on 24 inch side spacing with the outlet bank on 12 inch. The steam flow is parallel with the gas flow and the outlet tubes are located within the tube bank to shield them from high temperature radiation.

The main steam and hot reheat piping leads are each split into two half-capacity runs to increase flexibility. A single cold reheat lead is used. All these connect at the top of the boiler setting. Because of the long outdoor vertical runs, and the high design wind loadings, special structural lateral restraints were required.

Automatic extraction line shutoff, condensate and feedwater bypassing of heaters, turbine and extraction line drains, and large capacity heater emergency drains are used to prevent turbine water damage. These measures meet or exceed ASME recommended practices.

The heat cycle is basically a standard seven heater arrangement with an open dearating heater at the midpoint. The crossover point extraction (fifth heater) supplies steam to the two partial capacity boiler feed pump drive turbines and to the air preheater, the scrubber reheater coils, and to the closed Heater No. 5. A small motor driven boiler feed pump is used for start-up.

All pumps in the cycle except the main boiler feed pump are direct motor driven. Three partial capacity condensate pumps are used. All other pumps, blowers, and compressors serving the main cycle have complete 100% backup.

Heaters are in a single string and all closed heaters have stainless steel tubing.

The reheater is all pendant surface located behind the secondary superheater. There are three tube banks arranged for counterflow, with the reheater outlet tubes located within the bank for shielding.

Main steam temperature is controlled by spray attemperators while reheat temperature is maintained by gas recirculation with reheat spray attemperation available for upsets.

The economizer consists of 2" O.D in-line, bare tubes arranged in several banks such that a flue gas takeoff can be utilized to provide a higher gas temperature to the primary air heater.

The boiler ash cleaning system is by Diamond Power. The blowing medium is superheated steam. The system consists of 76 furnace wall blowers, 56 long retractable convection pass blowers and 24 extended lance blowers in the economizer banks.

F. AIR PREHEATERS

The primary air to the pulverizers and the secondary air to the furnace combustion zone is preheated by B&W/Rothemuhle Regnerative Air Heaters. For heating secondary air, two (2) vertical shaft 28-foot heaters are used, each with 185,000 square feet of surface arranged 64" deep. For primary air, one 35 foot vertical shaft heater is used with a surface depth of 70" and containing 314,000 square feet of surface.

A combination water wash and fire extinguishing system is located at the top of each air heater. The air heaters have special traversing blowers at the top and bottom of the heating elements.

G. FORCED DRAFT AND INDUCED DRAFT FANS

The combined power requirement of the forced draft, induced draft, and primary air fan system is about 13.2 MW, or about 3 percent of the

station output. TLT-Babcock variable pitch axial flow fans are used on all but the gas recirculation applications.

This unit is designed to burn a coal with a variety of characteristics, therefore, the fan requirements can vary considerably. One of the most valuable characteristics of the variable pitch axial flow fan is that it maintains a relatively high efficiency over a wide load range. A high fan efficiency, generally exceeding 80 percent can be maintained under most operating circumstances.

Axial flow I.D. fans have an important secondary benefit associated with furnace implosion prevention. Because of the stall characteristics of this fan, it does not produce high negative pressures at low flows. In contrast, a radial fan normally achieves its highest head at low flows. Under conditions normally associated with furnace implosion an axial I.D. fan would create only about one-half the negative perssure as compared to a radial fan.

H. CIRCULATING WATER SYSTEM

Water is recirculated from and back to the ash pond through the bottom ash and economizer hoppers continuously for cooling. Water for conveying of bottom ash, pyrites, and economizer ash, and operations of the fly ash vacuum system and dustless unloaders is pumped from the ash pond. The heat load from the ash, plus the high average natural evaporation rate, requires makeup of water to the pond. This makeup, plus that required for SO2 scrubber makeup, is taken from the cooling tower blowdown. This use results in cooling tower blowdown rates exceeding that which would be otherwise required for control of circulating water dissolved solids control.

Because of the possibility of high alkalinity in the ash water, acid can be fed at the ash pond. Further, the cooling tower blowdown is introduced at the ash water pump suction header to provide some dilution.

The cooling system for the turbine generator and its auxiliaries utilizes a single mechanical draft cooling tower. The tower is of wood construction and has eighteen cells. The surface condenser is two-pass 270,000 $ft.^2$, designed for 206,000 gpm of water at 87F. The heat exchange surface is composed of 1", 22 gauge type 304 stainless steel tubes, with 43'=8" effective length. Major plant auxiliary coolers are in parallel with the surface condenser and their cooling load is handled by the cooling tower. The heat load of the incoming well water (130F) for makeup is also handled by a smaller, separate cooling tower.

The main cooling tower is designed for maximum fire protection. It has fire-resistant fan stacks and fill, cement asbestos board cell bulkheads and other convering, plus a complete automatic sprinkler system.

A design cold water temperature of 87F was found to be optimum considering turbine performance, expected ambient temperature and humidity levels.

The demineralizer is sized for maximum expected soot blowing and boiler blowdown. Two full capacity trains, each having cation and anion exchangers, a degasifier and a mixed bed polishing unit are used. Demineralized water is stored in two 300,000 gallon tanks, which together represent 2-1/2 days of normal expected use for the unit. Raw water is taken directly from the well gathering system ahead of the storage pond and is normally free of turbidity and organic substances. The water is of a quality to permit use of only sand filters as pretreatment ahead of the demineralizer. Boiler continuous blowdown is recovered and directed to the demineralizer.

The demineralizer effluent is of sufficient quality to normally eliminate the need for a condensate polisher in the cycle. However, a condensate polisher is installed as insurance against excursions in feedwater quality. Control of pH and dissolved 02 is an all-volatile chemical feed system which monitors cycle and boiler water and steam quality continuously, while controlling the feed of the volatile chemicals automatically.

All demineralizer waste is sent to the ash pond.

I. ELECTROSTATIC PRECIPITATOR

Flue gas particulate control for this unit is provided by a B&W/Rothemuhle Electrostatic Precipitator System. The precipitator system consist of two casings, each 30' deep and 79' wide. Within each casing are four fields 14.5 feet deep and 42 feet high.

The precipitator has demonstrated a collection efficiency of 99.97 percent at an inlet loading of 20.4 grains per cubic foot of gas, with the outlet loading well below State and Federal requirements.

The rigid frame discharge electrodes have a maximum unsupported wire length of 7 feet which should minimize wire breakage. The rapper system for the collector plates and discharge frames is simple and ruggedly constructed and requires little maintenance.

The rappers rap individual collection curtains and frames on a staggered sequence, thereby reducing re-entrainment.

The electrostatic precipitator is dividied into two parallel flow chambers. Each of these chambers has four rows of hoppers with four hoppers per row (16 total). The front row of four hoppers in each chamber collect up to 67 tons of fly-ash per hour. The remaining 12 hoppers in each chamber collect up to 16 tons of fly-ash per hour.

J. FLUE GAS DESULFURIZATION

The flue gas desulfurization system supplied by Babcock & Wilcox consists of four 33 percent capacity limestone scrubber modules augmented by recirculation pumps, tanks, mixers, limestone mill, hot air reheat system, thickeners, piping and controls. This system was designed for 86 percent efficiency with flue gas containing up to 4400 ppm SO2 using limestone containing 95 percent calcium carbonate. Actual composition of the inlet flue gas has ranged between 2800-3300 ppm SO2.

Base loading of the San Miguel Plant will involve the daily consumption of over 600 tons of limestone delivered to the plant site by rail.

The flue gas is directed to the scrubber modules by the two induced draft axial flow fans. No booster fans are required. The scrubber modules are arranged off a common inlet plenum behind the stack and any number of modules can be isolated from the flue gas path by closing module inlet and outlet isolation dampers. There is also bypass capability so that when SO2 emissions are within required limits, hot flue gas can be diverted to the exit plenum to minimize reheat costs and its effect on turbine output. Flue gas enters a module through a down flow quencher section. Each quencher has a 350 hp recirculation pump which recirculates limestone slurry to a degritting hydrocyclone and then to the quencher spray and wall wash nozzles. The flue gas leaving the quencher makes a 180 degree turn and enters a two stage countercurrent tray absorber. Two 400 hp absorber pumps per module recirculate limestone slurry to the high capacity, low back pressure spray nozzles.

The chemistry for this system is maintained by continuous pH monitoring of both the quencher and absorber recirculation tanks. In this dual loop/plug flow slurry recirculation system, differentiated chemistry conditions are maintained for each component of the scrubber module to insure maximum sulfur dioxide removal while minimizing the excess consumption of limestone. Plug flow is incorporated by baffling each tank in a manner that requires the maximum in slurry residence time, thus promoting higher calcium carbonate dissolution and product oxidation in the tank.

Flue gas leaving the absorber section passes through a wash tray and a two-stage moisture separator section before entering a common reheat mix chamber located in the exit plenum approaching the stack inlet. Flue gas reheating is accomplished by mixing a stream of ambient air, heated by extracted steam to 400°F, with the flue gas by using two parallel pressure saving air foils with side ports. The combined gas and air mixture temperature is raised to a temperature 30-40°F above the dew point.

The 3/4"X 3/8" limestone is ground to 95 percent minus 325 mesh fineness in a wet ball milling system capable of handling 48 tons per hour. The limestone feed slurry is stored in a 5 hour capacity slurry storage tank. The 30 percent solids pulverized limestone feed slurry is pumped to the module recirculation tanks to react with the sulfur oxides absorbed in the module. The original 3/4" X 0" limestone specification had to be changed since it was not possible to handle this material when damp. Severe pluggage in all parts of the system caused high maintenance costs and operating problems.

Spent products from the scrubber system are removed with a bleed stream and directed to two 115 foot diameter thickeners. Clarified water from the thickener overflow is recycled to the scrubber and milling system for reuse. Water from the ash pond is used as make-up as required. A portion of the recycle water is blended with cooling tower blow-down water for use as the moisture separator wash. The thickener sludge is dewaterd and mixed with fly ash and then trucked to the mine for disposal.

The operating electric power requirement for the sulfur dioxide removal system is 7,900 KW, or 1.8 percent of the gross generator output. The average reheat steam requirement is 113, 700 lb/hr of steam at 179 psi and 718°F.

K. STACK

The 450' chimney with 25' diameter top and 40' diameter bottom is constructed of reinforced concrete and has a chemical resistant brick liner conforming to ASTM-C279, Type H. A potassium silicate base mortar that is completely inorganic is utilized in construction of the brick liner.

Sampling ports are placed to conform to Texas Air Quality Board requirements. Access platforms and high intensity obstruction lights are installed to conform to OSHA and FFA requirements, respectively. A 900 pound personnel elevator is installed from ground level up to the stack monitoring level.

L. LIGNITE HANDLING SYSTEM

Base loading of the San Miguel Plant will involve the daily consumption of approximately 10,000 tons of lignite.

The run-of-mine lignite is received at the plant site from the adjacent surface mine in both 50-ton and 100-ton capacity dump trucks. Based on hauling five days per week and two 8-hour shifts per day, truck deliveries can occur at a rate of 14 trucks per hour, depending on the quality of the fuel. The trucks dump directly into a 450-ton capacity hopper, from which the lignite

is fed via vibrating feeders onto an inclined conveyor rated at 1400 tons per hour. The coarse material is delivered to a rotary breaker and the fines to a reversible conveyor under the rotary breaker. The rotary breaker reduces the lignite to 6" X 0" and discharges on the reversible conveyor. The reversible conveyor delivers all of the lignite to either a conveyor feeding the ready storage pile or to a conveyor feeding two 70' diameter, 200' high concrete silos for live storage. Both of these latter conveyors are equipped with belt scales for continuous weighing and recording of the lignite deliveries.

Lignite is recovered from the concrete silos by one of two rotary plows, which operate in tandem, and/or from the live storage pile by vibrating feeders and conveyed to a surge bin at the crusher house where two ring-type granulators will further reduce the size to 3/4" X 0". The crushed lignite is then conveyed to a transfer tower surge bin and then to tripper conveyors which feed the seven coal bunkers.

The two concrete silos have a combined net storage capacity of 28,000 tons, sufficient for a normal weekend supply of fuel. The ready storage pile normally contains 14,000 tons, sufficient for at least one day supply of fuel. The dead storage area contains sufficient reserves to allow operation for 40-60 days.

There are several features incorporated in this system which were specifically designed to handle a non free-flowing fuel with high spontaneous combustion tendencies.

The use of silos results in achieving a first-in, first-out concept which is necessary with fuel which has high spontaneous combustion tendencies. Slotted openings and "parabolic" hoppers in the bottom of the silos should prevent arching. The use of rotary plows instead of vibrating feeders to remove material from the silos should further insure against arching. The use of rotary plows underneath concrete silos is the first installation of this kind in the United States.

Special attention was given throughout the system to eliminate flow problems. Hopper angles are steep. Stainless steel liners are used in hoppers, chutes and feeders. Low conveyor speeds are utilized.

Solid state electronics and programmable controllers are used to obtain complete automation. The system is designed to obtain maximum flexibility of operation.

M. LIMESTONE HANDLING SYSTEM

The flue gas desulfurization system requires approximately

40 tons of limestone per hour.

Limestone delivered in 100-ton railroad cars is dumped into a 300-ton hopper. An inclined conveyor elevates the limestone to a tripper conveyor installed in the top of a closed A-frame building. This building has 375,000 cu. ft. of storage capacity.

The storage building has a steep vee-shaped floor under which a low-profile single opening rotary plow removes limestone from the building and places it onto a horizontal conveyor. The horizontal conveyor transports the limestone to an inclined conveyor which in turn delivers limestone to a storage bin ahead of the wet limestone ball mill. Refilling of the bin is accomplished at least once each eight-hour shift.

N. ASH DISPOSAL SYSTEM

Separate systems for removal of fly ash and bottom ash are utilized. The bottom ash system removes ash from the furnace bottom, economizer hopper and pyrites hoppers. The fly ash system removes ash from the precipitator hoppers and the gas recirculation fan dust collection hoppers. The total ash production can equal 150 tons per hour.

Bottom ash is collected in a three compartment, water impounded, 13,000 cu. ft. hopper. Each compartment is equipped with two discharge doors, one on each side of the compartment. These doors have extra large, 6.25 $ft.^2$ openings. Each discharge door enclosure is equipped with a jet pump and crusher. The discharges from the crushers are piped together on each side to form two parallel systems. Four hundred tons per hour of bottom ash can be pumped hydraulically from either of these parallel systems to either of two 45 ft. diameter, 60,000 cu. ft. dewatering bins or directly to the ash pond.

Pyrite rejects from the lignite pulverizers are conveyed hydraulically from the pulverizers to an 1800 cu. ft. storage tank which is periodically discharged through either of the bottom ash conveyor lines.

The ash collected in the four economizer hoppers continuously downspouts into two water filled collector tanks. Ash stored in these two tanks is periodically sluiced to the pyrites storage tank for removal. Provisions have been made for future installation of clinker grinders should they be required.

Two separate pressurized fly ash conveying systems are utilized, one for each compartment's front row of hoppers. A common vacuum conveying system is utilized to remove fly ash from the remaining twelve hoppers of both compartments.

Each pressurized fly ash system also removes ash from two of the four hoppers of the gas recirculation system dust collectors.

The three fly ash systems operate approximately 50% of the time, simultaneously. These systems convey the ash to a 148,000 cu. ft. steel storage bin for disposal. This bin can be by-passed, with all fly ash directed to the ash ponds.

The bottom, economizer, and air heater ash from the dewatering bins is loaded into trucks and used in maintaining the haul roads. Any surplus of these ashes is disposed of with the fly ash.

The fly ash is either conditioned with water to make it truckable and returned to the mine, mixed with dewatered slurry from the SO2 removal system, or sold to various users.

O. PLANT MANAGEMENT

Automation of plant components and systems is utilized to a great extent. Boiler and scrubber controls are covered under the description of the boiler. The soot blowing system has a microprocessor based control system. The turbine generator has a Westinghouse digital electrohydraulic governor system, Model II, using their Prodac 2500 computer. The lignite and limestone handling systems use programmable logic controllers. The ash handling system is automated to the maximum practical extent. Solid state integrated logic control systems are used for cooling tower and ash water systems. The makeup demineralizer plant is fully automatic.

Data gathering and processing, monitoring and alarm of all plant systems is done by a Honeywell 4500 computer system which presents useful information to the plant operators by means of color CRT's, alarm printers and logging typers. The computer also has so-called "smart" remote terminals, have redundant CPU and sufficient memory and storage to handle plant historical data, environmental data, analyze plant performance and provide for post-trip critique.

The boiler analog control system is a Bailey 820 Electric Analog system which acts in coordination with the pulverizer/burner monitoring and management system.

Modulating operators for dampers are elecrtic, with pneumatic pistons used where only on/off or open/shut type of operation is needed. Control valves are pneumatic with 4-20 ma current input positioners.

The system includes an integrated boiler-turbine master control, which controls the seven pulverizers, two primary air fans, two F.D. fans and two I.D. fans. Coordinated with the combustion control system is the feedwater control system which controlls the two variable speed boiler feed pumps, and the superheat and reheat control system which controls the flue gas recirculation fan and spray attemperators.

A Bailey 762 Solid State Digital Sequence Logic System monitors and manages the start-up and shut-down of the seven pulverizers and forty-two burners in the seven compartments. The action of this system is coordinated with the combustion control system through the sending and receiving of permissives. Through these interconnections, the systems perform as one total system.

The operator interface consistes of an integrated control station for each pulverizer coordinating all the operator functions as they relate to a particular pulverizer. This one station per pulverizer contains all the Manual/Auto functions of a conventional analog system plus all the operator used devices for the Pulverizer/Burner system. In addition, the station includes a mimic arrangement of the pulverizer and compartment with LED status of all dampers, burners, etc., plus status of the "Start" and "Stop" sequence by means of LED sequence display.

The scrubber control system consists of multiple simple loops, and is implemented with the Bailey 7000 unitized electronic analog control. The equipment incorporates both the controller and the manual automatic station used by the operator into a single panel mounted device. Control valves are pneumatic with a 4-20 ma input to the valve positioners. Control systems are supplied for four scrubber modules, two thickeners, and two reheaters. A stack by-pass control protects against overpressure.

P. SUBSTATIONS

Output from the generator is routed by an isolated phase bus to the 430 MVA, 22.8/345 KV unit step-up transformer and the 54 MVA,24/7.2 KV dual secondary unit auxiliary transformer.

The plant substation contains 345 KV and 138 KV sections interconnected by two 300 MVA autotransformers. The tertiary of each autotransformer is connected to dry-type shunt reactors to provide reactance to the 345 KV transmission system.

The classical breaker-and-a-half scheme is used in both the 345 KV and 138 KV sections of the substation. There are two 345 KV and four 138 KV outgoing lines plus one 138 KV line to the mine substations.

One 54 MVA, 138/7.2 KV dual secondary station reserve auxiliary transformer is installed. Dual secondaries are utilized on the unit auxiliary and the station reserve auxiliary transformers to limit fault current to the 7.2 KV switchgear. A nonsegregated phase bus connects the auxiliary transformers to the 7.2 KV switchgear.

Motors 500 hp and above operate at 6.9 KV. Power centers rated at 3 MVA, 7.2/.48 KV are utilized to serve motors below 500 hp. There are eight complete power centers located near the load centers of the plant. Standard motor control centers are utilized. Primary lighting is at 277 volts.

Critical service needs for items such as the booster fire pumps, inverter, battery charger, burner controls, turbine controls and plant computer systems are served by a critical service system utilizing an automatic transfer switch between the normal power supply and the Emergency Diesel Generator.

Q. FIRE PROTECTION

The basic fire protection scheme is typical for large coal fired plants and includes the following: (1) engine driven pumps at the storage pond, backed up with motor driven pumps at the cooling tower basin; (2) multiple automatic sprinkler systems; (3) looped firemain with fire plugs at 300 foot intervals (200 feet at boiler and coal storage) and (4) Halon systems in electronics and control rooms. One unusual feature is the need for a booster system similar to that for high rise buildings to serve the sprinklers and hose stations for protection of the upper lignite conveyors, transfer points, silos and bins.

R. SERVICE FACILITIES

The main control room is located on the turbine deck. The boiler, turbine-generator, lignite handling system, and major components of the power plant accessories are controlled from this point. An auxiliary control room located at ground level contains the controls for the electrostatic precipitator, SO2 scrubber system, limestone handling, ash handling, and ash water systems, plus the plant's water and fuel quality laboratory.

Several other structrues, including a combined office and maintenance building, warehouses, the lignite handling system buildings, and a limestone storage building are located on the plant site.

S. MINING

An independent contractor operates the mine with responsibility for delivering the lignite to the truck hopper at the plant site. A Marion walking dragline with a 65 cubic yard bucket and a Page walking dragline with a 30 cubic yard bucket belonging to San Miguel are used by the contractor. All other mining equipment and trucks are provided by the mining contractor. Additional facilities at the mine furnished by San Miguel include office, warehouse and maintenance buildings. The mine is located approximately two miles southeast of the plant site and is connected to the plant site by a 150 ft. all-weather haul road. Electric power is supplied to the mine facilities via a 138 KV line from the plant substation.

San Miguel has under lease approximately 22,000 acres of land with fuel reserves estimated to last about 75 years. The Lower Jackson-Yegua lignite occurs in four seams with the top seam about 40 feet below ground level. The top three seams are about one foot thick, the lower seam is about 2-3 feet thick with the parting between each of the seams about one foot thick. Careful attention paid to mining these thin seams with Easi-Miners has resulted in a consistently higher quality of fuel than that used during the design phase. The high moisture content of the lignite causes handling problems, but simplifies dust suppression.

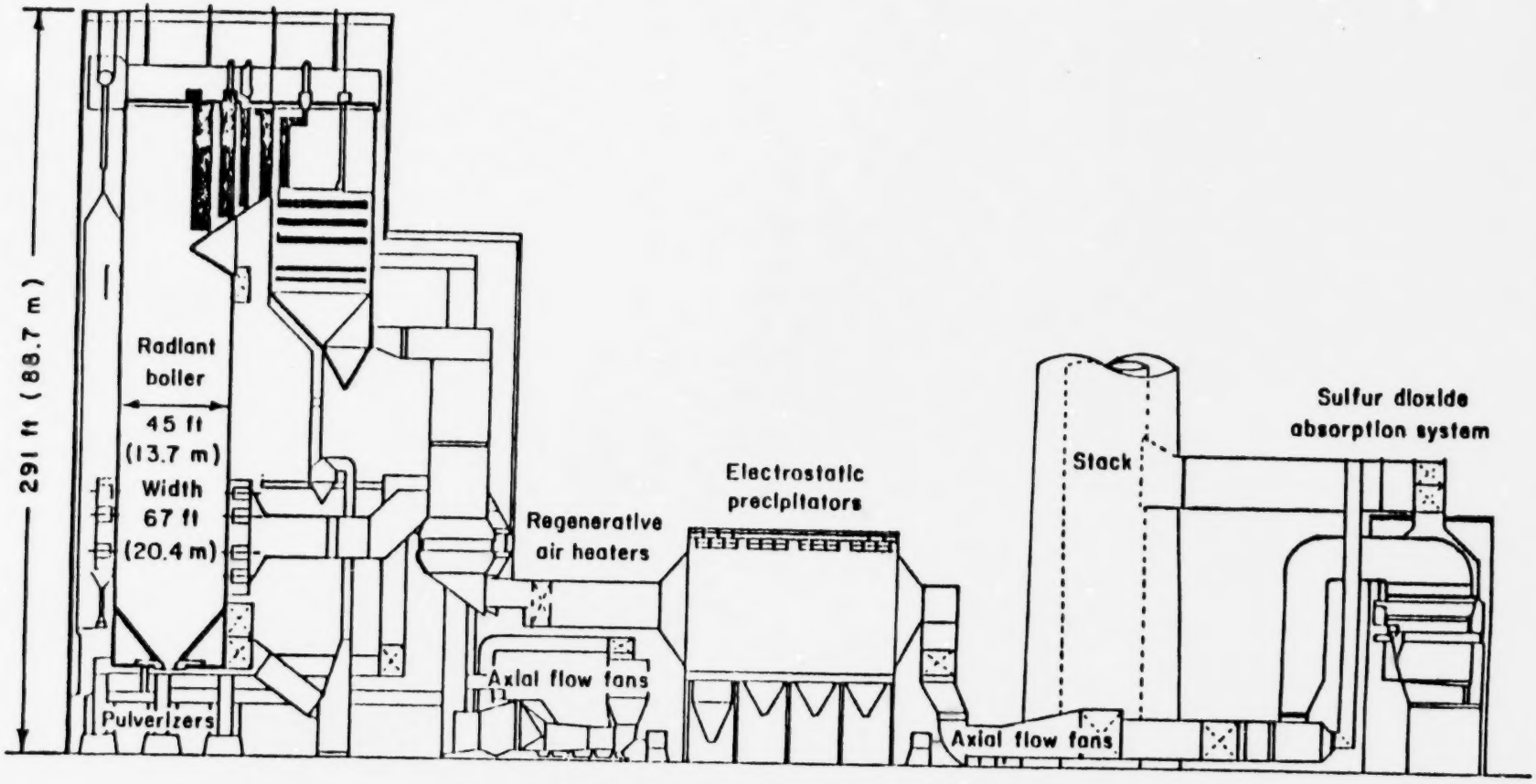


Figure 1. Cross-sectional view of 400-MW pulverized-coal-fired San Miguel Plant, Christine, Texas. Designed to burn a high-ash Texas lignite.

Source: Gronhovd, G. H. and E. A. Sondreal, "Technology and Use of Low-Rank Coals in the U.S.A.", presented at the Seminar on Technologies for the Utilization of Low Calorie Fuels, Varna, Bulgaria, April 20-22, 1976.

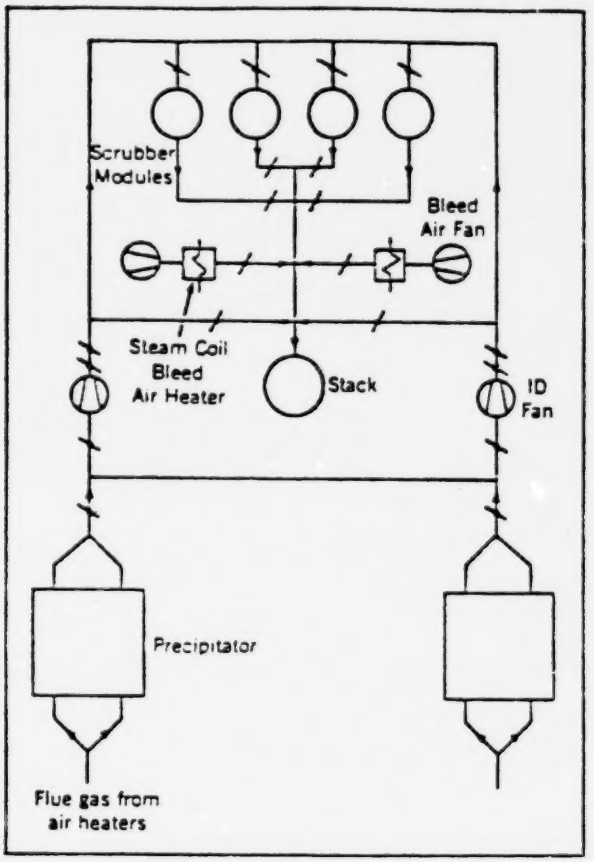


Figure 2. Emissions control equipment (precipitators and SO_2 removal system)

Source: Haller, K. H., "Design for Lignite Firing at the San Miguel Power Plant", presented at the 11th Biennial Lignite Symposium, San Antonio, Texas, June 15-17, 1981.

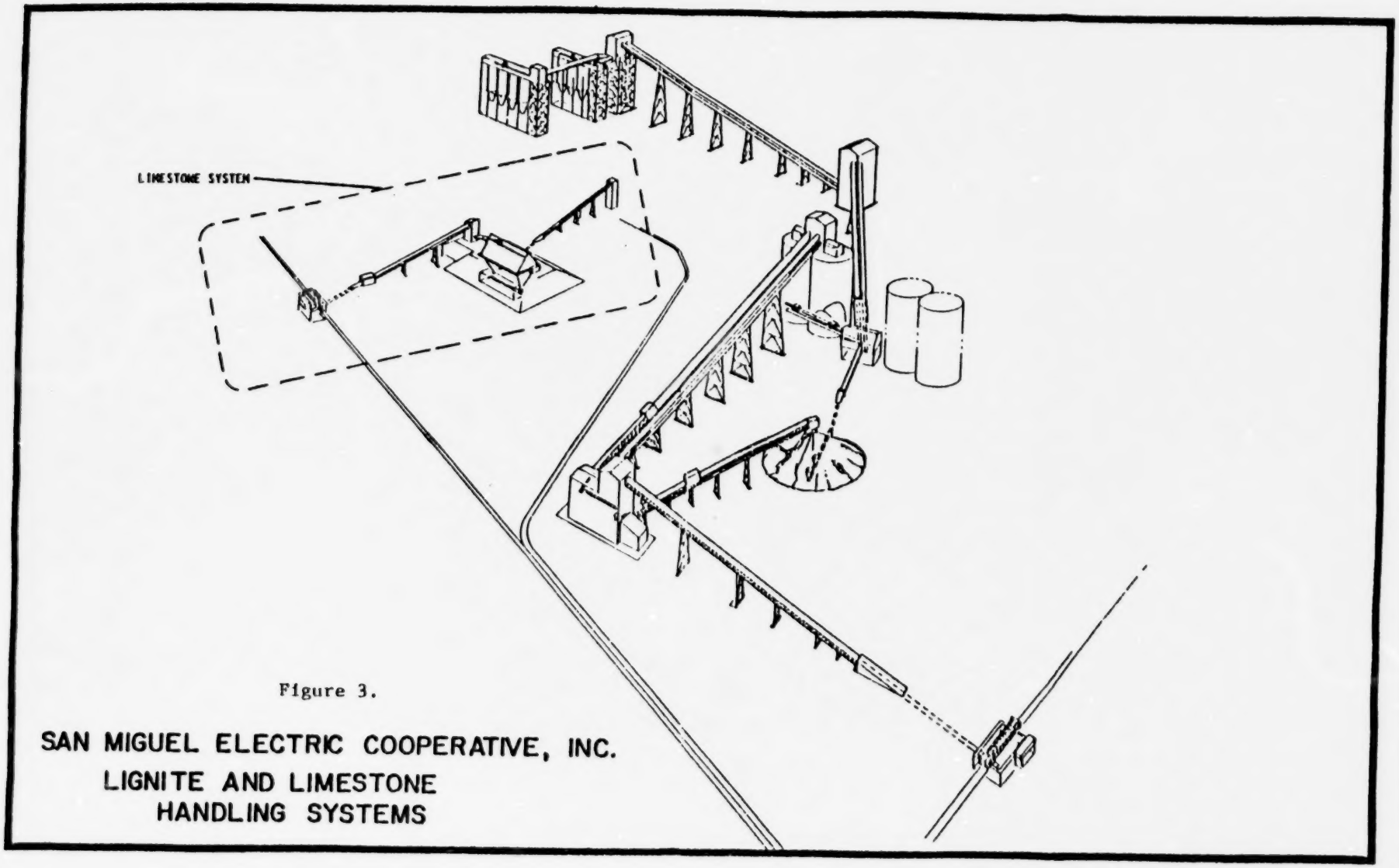


Figure 3.

SAN MIGUEL ELECTRIC COOPERATIVE, INC.
LIGNITE AND LIMESTONE
HANDLING SYSTEMS

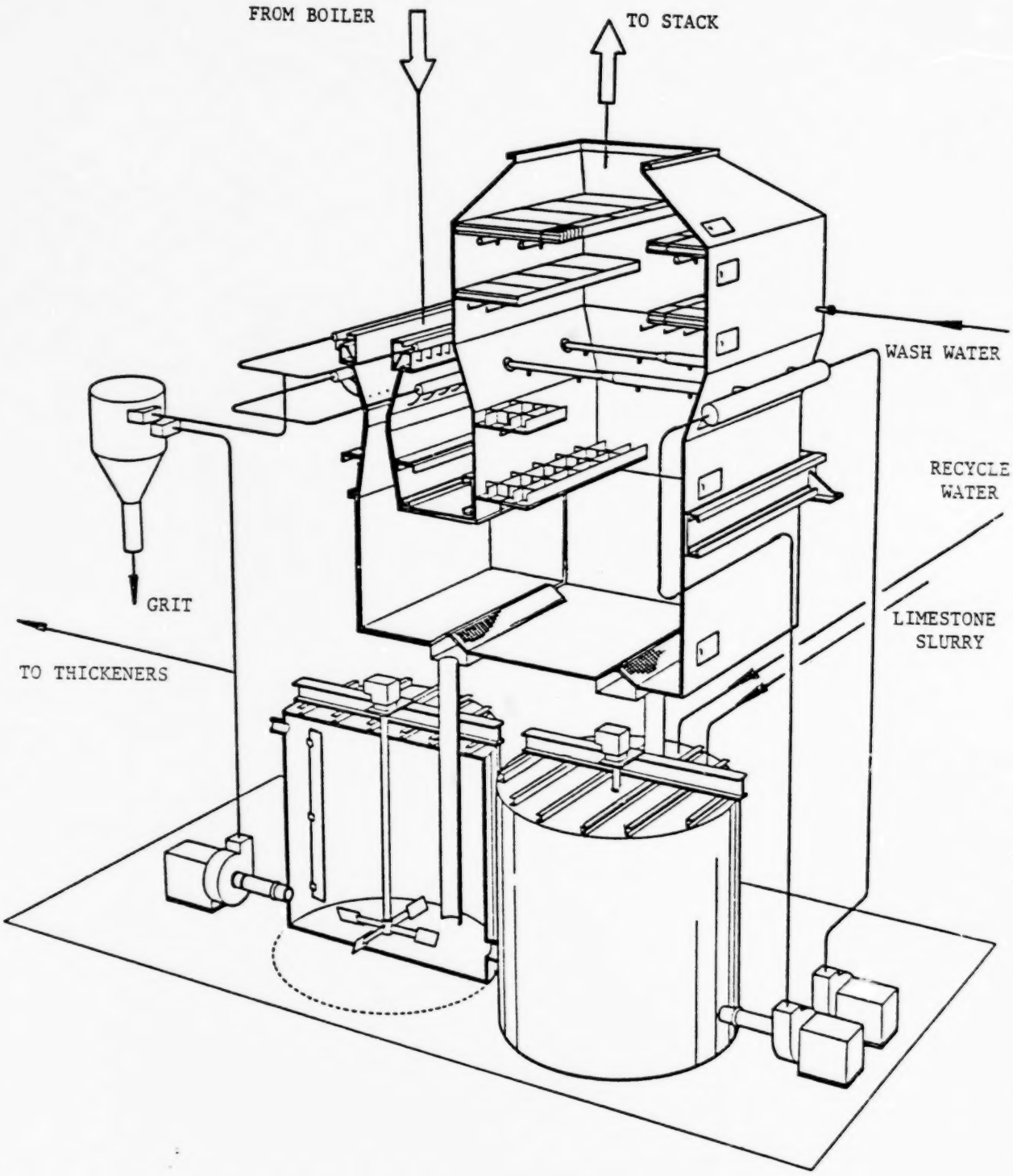
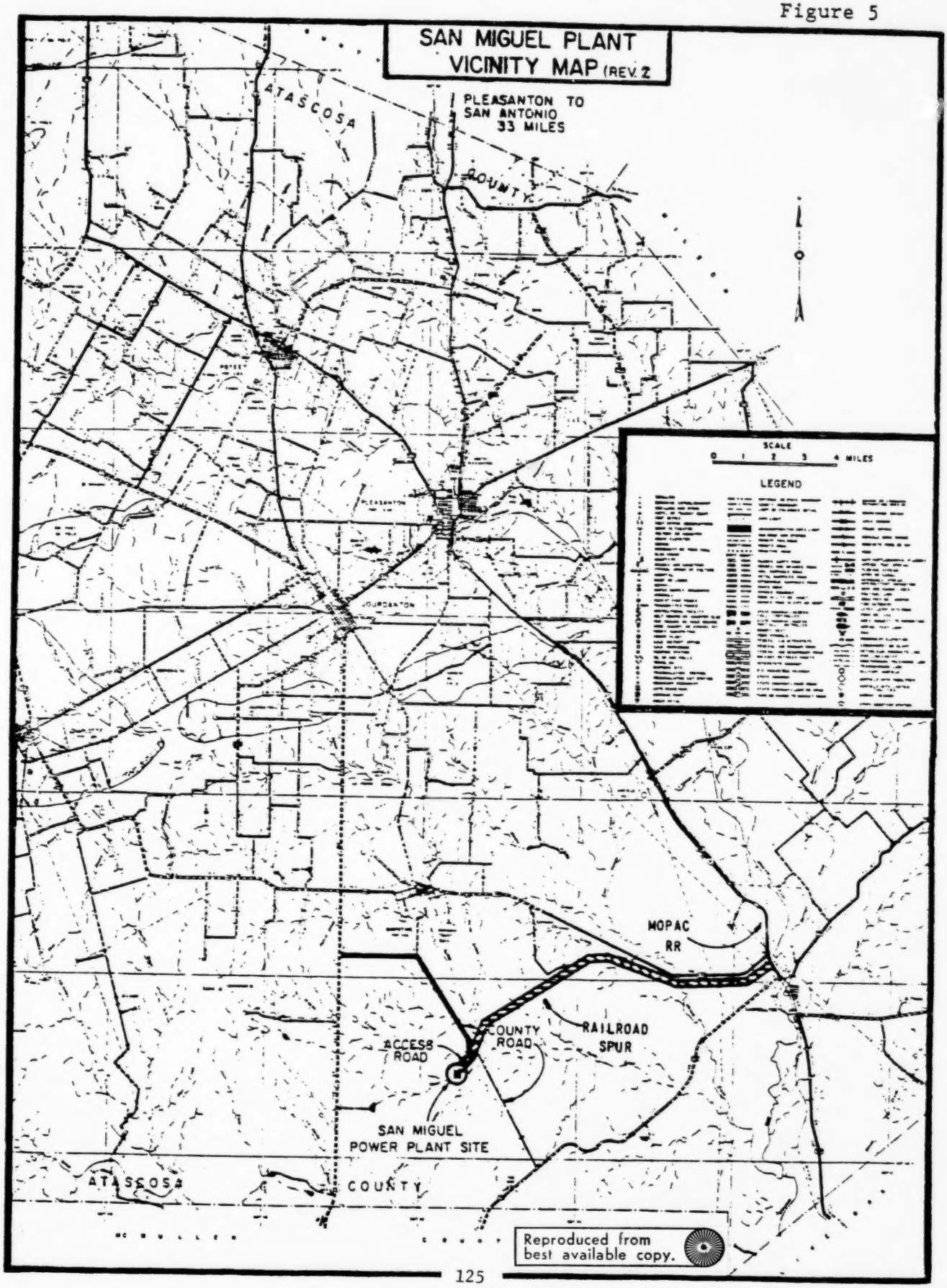


Figure 4. FLUE GAS DESULFURIZATION MODULE

Figure 5



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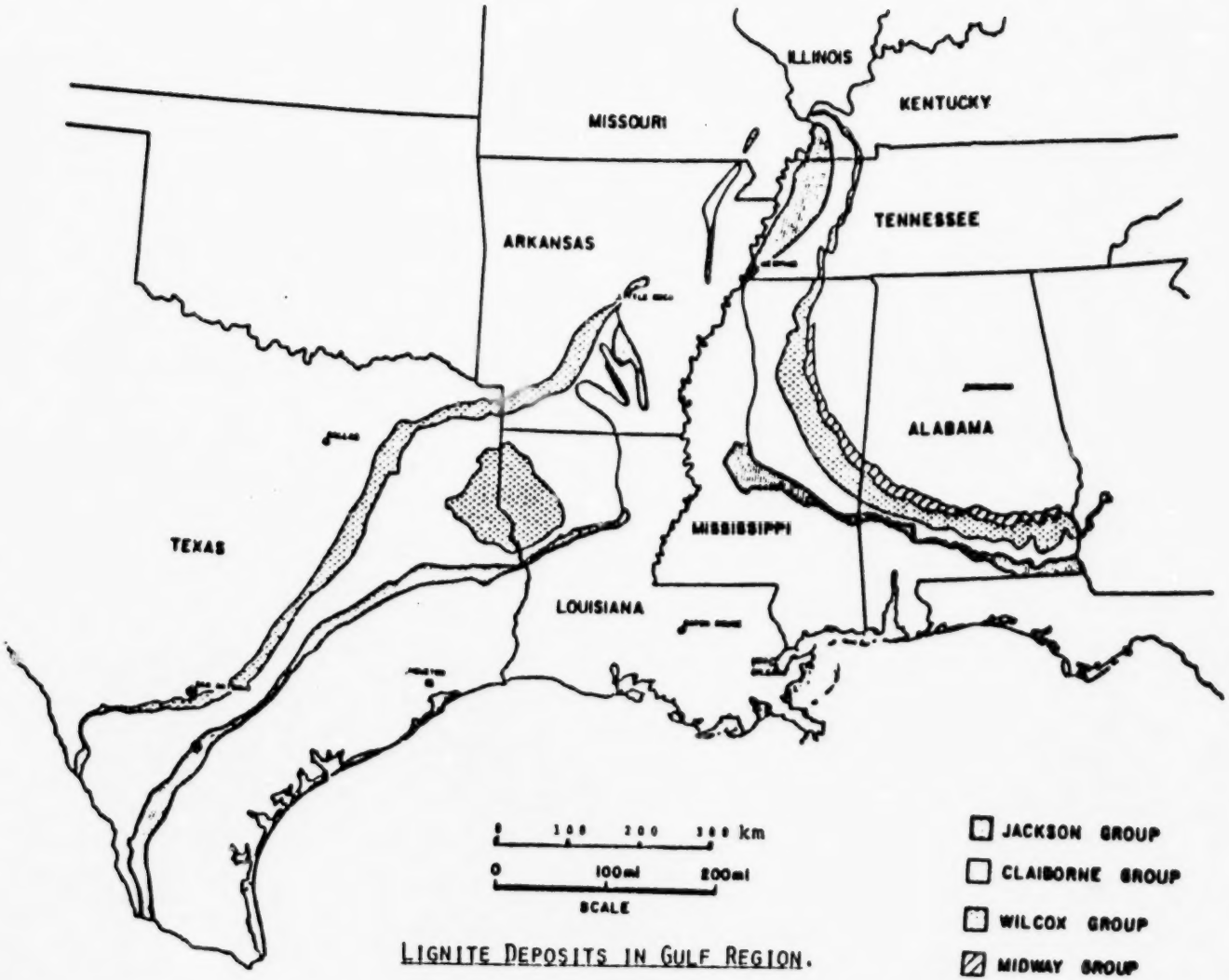

LIGNITE DEPOSITS IN GULF REGION.

Figure 6.

Table 1. Boiler and Equipment Specifications for San Miguel Plant

Plant name	San Miguel	
Plant rating, MW	400	
Boiler specifications:		
Manufacturer	Babcock and Wilcox	
Firing method	pulverized coal	
Rated capacity, lb/hr	$3,054 \times 10^3$	
Pressure, lb/sq.in. gauge	2,620 (178 atm)	
Steam temp. °F (°C)	1,005 (541)	
Reheat temp. °F (°C)	1,005 (541)	
Balanced draft furnace	Yes	
Heat release at rated capacity:		
Volume heat release, Btu/hr/ft^3 (kcal/hr/m^3)	7,500 (66,744)	
Area heat release, Btu/hr/ft^2 (kcal/hr/m^2)	85,000 (230,562)	
Furnace height, ft (m)	220 (57.1)	
Furnace width, ft (m)	67 (13.7)	
Furnace depth, ft (m)	45 (20.4)	
Design flue gas temp. and tube spacing:	Temp. °F (°C)	Tube spacing center-to-center inches (cm)
Entering secondary superheater	1850 (1010)	24 (60.9)
Entering reheater	1500 (815)	9 (22.9)
Entering primary superheater	1110 (598)	4.5 (11.4)
Entering economizer	825 (440)	4.5 (11.4)
Soot blowers:		
Type	Steam	
Number of short wall blowers	76	
Number of long retractable blowers	80	
Pulverizer, burner data:		
Pulverizer model	Babcock and Wilcox - MPS 89	
Capacity, tons/hr	86	
Number per boiler	7	
Number of burners per boiler	42	

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TABLE NO. 2
BASIS FOR DESIGN
ATASCOSA-McMULLEN LIGNITE

	Average Fuel Analysis				
	As Received Basis	Moisture Free Basis	Moisture & Ash Free Basis	Best Fuel As Received	Worst Fuel As Received
Moisture Total, %	30.00	-	-	27.0	35.0
Moisture Equilibrium, %	29.00	-	-	28.0	32.0
Volatile Matter, %	23.20	33.20	55.90	-	-
Fixed Carbon, %	18.40	26.20	44.10	-	-
Ash, %	28.40	40.60	-	21.0	29.0
Carbon, %	29.05	41.50	69.86	-	-
Hydrogen, %	2.38	3.40	5.73	-	-
Nitrogen, %	0.51	0.73	1.23	-	-
Sulfur, %	1.67	2.39	4.02	-	-
Chlorine, %	0.06	0.08	0.13	-	-
Oxygen, %	7.93	11.30	19.03	-	-
Btu/lb.	5,000	7,143	12,025	6,350	4,200
Sulfur Pyritic, %	0.45	0.65	-	-	-
HGI	-	92	-	-	-

Ash Analysis, %	
SiO_2	63.0
Al_2O_3	17.5
TiO_2	0.8
Fe_2O_3	2.8
CaO	4.8
MgO	0.7
Na_2O	3.1
K_2O	2.0
SO_3	4.6
P_2O_5	0.1
Undetermined	0.6

Ash Fusibility, °F	Reducing	Oxidizing
IDT	2120	2190
SST	2380	2430
HST	2430	2480
FT	2700+	2700+

Table 3. Proximate and Ultimate Coal Analyses

(As Received)

Date Sampled	5/25/82	5/26/82	5/27/82	Average
Proximate				
% Moisture	28.76	27.43	30.85	29.01
% Ash	26.81	26.73	25.99	26.51
% Volatile	25.60	25.30	24.86	25.25
% Fixed Carbon	18.83	20.54	18.30	19.22
BTU/lb	5718	5971	5426	5705
% Sulfur	1.71	1.80	1.90	1.80
Ultimate				
% Moisture	28.76	27.43	30.85	29.01
% Carbon	32.03	32.80	29.63	31.49
% Hydrogen	2.66	2.84	2.43	2.64
% Nitrogen	0.72	0.72	1.08	0.84
% Chlorine	0.31	0.40	0.35	0.35
% Sulfur	1.71	1.80	1.90	1.80
% Ash	26.81	26.73	25.99	26.51
% Oxygen	7.00	7.28	7.77	7.35

Table 4. Chemical Analysis of Fly Ash Samples[+]

	Inlet ESP Hopper 5-26-82	Mass Train Thimbles 38 MT Run #4	44 MT Run #5	48 MT Run #3	49 MT Run #2
Li_2O	0.02	0.02			
Na_2O	3.9	3.8	3.9	3.7	3.5
K_2O	1.6	1.5			
MgO	0.67	0.65			
CaO	4.6	4.5			
Fe_2O_3	3.8	3.8			
Al_2O_3	21.9	22.4			
SiO_2	62.2	61.9			
TiO_2	1.7	1.7			
P_2O_5	0.05	0.05			
SO_3	1.2	0.62			
LOI	0.3	<0.03	0.09	<0.03	<0.03
$SO_4^=$	0.41	0.38			

[+]Analyses done on a digested sample of fly ash which had been ignited at 750°C.

APPENDIX C

LIST OF REGISTRANTS

EPA's Industry Briefing on the
Organic-Acid-Enhanced Limestone FGD Process

LIST OF REGISTRANTS

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